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Boiler Design Criteria for Dry Sorbent SO2 Control with Low-NOx Burners

Combustion Engineering, Inc., Windsor, CT

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BOILER DESIGN CRITERIA FOR DRY SORBENT SO_2 CONTROL WITH LOW-NO_x BURNERS

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Coal	Limestone	Stationary Sources	21D	08G
Boilers		Dry Sorbents	13A	
Design		Low-NOx Burners	14G	
Sulfur Dioxide		LIMB	07B	
Nitrogen Oxides				

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ABSTRACT

A program to develop boiler design criteria for application of dry sorbent control technology with low-NO_x burners on tangentially fired pulverized-coal-burning boilers was conducted under EPA sponsorship. A comprehensive review of past and current research in the area of sorbent SO_x control was performed to provide a basis for evaluating the implications of this technology on boiler design, cost effectiveness, and operability. Historical and projected design trends were analyzed for all tangentially fired pulverized coal utility boilers built by C-E during the period 1960 to the present, including the effect of coal rank. A candidate host unit was selected for consideration as a site for demonstration of dry sorbent SO_2 control. Dry sorbent process designs, including sorbent preparation/delivery equipment and boiler modifications, were developed and costed for new and retrofit (200 MWe, 400 MWe, and 600 MWe) high-sulfur coal-fired units and new (200 MWe, 400 MWe, and 600 MWe) low-sulfur coal-fired units. SO_2 removed was 50% in the boiler for all cases. Spray dryers were incorporated on the new units to achieve overall SO_2 removal (sorbent injection plus spray dryer) of 70% (low-sulfur coal) and 90% (high-sulfur coal). Conventional limestone flue gas desulfurization costs were developed for comparison purposes. Capital costs, cost of electricity, and cost effectiveness per ton of SO_2 removed were developed.

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SECTION 1

INTRODUCTION

Increasing concern for the environmental effects of acid rain has led to a re-examination of the fossil fuel burning electric utility industry. Previously, only new plants were mandated to install pollution control equipment for gaseous emissions under federal New Source Performance Standards (NSPS), although some existing plants were retrofitted with equipment due to local requirements. However, proposed legislation in response to the acid rain question would require at least a portion of existing oil and coal burning plants to limit emissions of sulfur dioxide (SO_2).

Due to the possibility of regulation of existing plants, various alternative sulfur oxide removal processes are being examined in addition to the tail end lime or limestone scrubbing systems predominantly employed to meet NSPS. Because of the combined requirements of "retrofitability" and moderate sulfur removal efficiency, the concept of furnace sorbent injection has received renewed interest and study. In the early 1960's, furnace injection originated as a method for controlling SO_2 emissions without involving utilities in sophisticated chemical flue gas treatment systems. However, trials in small- scale furnaces and full-scale utility boilers generally failed to demonstrate sufficient in-furnace SO_2 removal at reasonable sorbent-to-sulfur ratios.

In retrospect, it appears that the high temperatures and oxidizing conditions typical of conventional utility boilers created conditions unfavorable for SO_2 absorption. The development of advanced low NO_x utility boiler combustion technologies may provide new combustion conditions and lower furnace gas temperatures which may be more suited for in-furnace absorption of SO_2 by limestone or other sorbents. The combination of these two technologies has been given the acronym LIMB, representing Limestone Injection with Multi-stage Burners for NO_x control.

This report presents the results of the EPA contract entitled "Boiler Design Criteria for Dry Sorbent SO_2 Control with Low-NO_x Burners". This study comprised several tasks: (1) literature survey of state-of-the-art knowledge of SO_2 removal by sorbent injection; (2) review of historical and projected design trends for C-E coal-fired utility boilers; (3) selection of a candidate host site for LIMB demonstration; (4) design of sorbent preparation systems plus definition of boiler and back end modifications required to incorporate LIMB on new and retrofit units; and, (5) economic analyses of these LIMB-related designs, plus, for new units, SO_2 removal via conventional flue gas desulfurization techniques.

SECTION 2

CONCLUSIONS

While early field tests of sorbent injection for SO_2 control were inconclusive, recent tests have produced more promising results. SO_2 removals of 40% to 50% have been achieved at Ca/S molar ratios of 2:1 to 5:1. It was concluded that the most significant negative factor in the early tests was the high local temperatures in the mixing region which caused deadburning of the injected sorbent.

The key factors which most directly determine the effectiveness of sorbent injection on a particular boiler are:

Coal type - Ash compositions vary considerably. Laboratory tests have shown that sorbent injection affects the slagging and fouling behavior of some coals. Ash reactivity can affect sorbent capture.

Sorbent type - $Ca(OH)_2$ was determined to be the most effective sorbent, followed by limestone. The most important physical characteristics are specific surface area and porosity which determine the calcination and sintering rates. Limestone was shown to have a maximum specific surface area at 1832°F. CaO absorption falls off sharply above 2192°F while $CaSO_4$ starts to decompose at 2200°F.

Time/temperature effects - SO_2 absorption is a strong function of the residence time within a critical temperature range. For limestone, this range was found to be roughly 2200°F to 1700°F. The point at which the sorbent is injected into the boiler is critical in maximizing the residence time within this temperature range and minimizing the exposure to higher temperatures which adversely affect sorbent performance. Maximum SO_2 absorption was obtained when the sorbent was injected in the area of the overfire air ports.

Stoichiometry - SO_2 absorption increases with increasing Ca/S molar ratios. Ca/S ratios above roughly 4:1 were shown to give only small improvements in SO_2 removal.

Gas composition - A reducing atmosphere in the reaction zone does not affect SO_2 absorption. The rate of SO_2 absorption is higher for higher concentration of SO_2 in the gas stream.

Existing ash collection systems may be adversely affected by sorbent injection. Electrostatic precipitator performance was shown to drop off considerably due to the change in particulate resistivity. Wet ash conveying systems were susceptable to plugging. No adverse effects were identified in the performance of mechanical collectors. Operating dry scrubber systems in conjunction with sorbent injection may offer economic advantages in some cases by allowing units firing high sulfur coals to meet NSPS limits.

For 50% SO_2 removal on either new or retrofit boilers, no changes to the boiler shell would be required. Economizer modifications would be required for the retrofit units. Some surface rearrangement would be required for the new units. For higher SO_2 removals with LIMB, the boiler design modifications would be more extensive.

Sorbent injection via dilute phase transport into the upper furnace area will require multiple injection points (typically 8 to 16 nozzles) with injection velocities of about 200 ft/sec. Based on theoretical analyses, it is not clear whether adequate penetration can be achieved with conventional injection approaches.

Back end upgrade via multiclone plus SO_3 conditioning appears to be preferable to ESP upgrades for retrofit units, based on capital cost requirements.

For retrofit units, incorporation of LIMB for 50% SO_2 removal costs 57.5 \$/kw at 560 MWe. Adding LNCFS burners for NO_x control adds another 2.27 \$/kw.

For new units at 600 MWe, upgrading for LIMB capability costs 93.1 \$/kw for high sulfur coal and 44.1 \$/kw for low-sulfur coal. This compares to 221.3 \$/kw and 151.1 \$/kw, respectively, for SO_2 removal via wet scrubbers.

SECTION 3

PROCESS DESCRIPTION AND EVOLUTION

Dry injection of alkaline earth compounds into utility boilers has been discussed and tested for the past 30 years. Initially, the primary reasons for injection involved boiler corrosion. In the 1950's, Combustion Engineering, Ref. (1), employed a commercial procedure for injection of magnesium oxide (MgO) to control molten phase, vanadium-related high temperature oil ash corrosion and SO_3 related cold-end corrosion and acid smuts fallout. In 1955, a patent issued to John W. Beckman covered a process to remove SO_2 and sulfur trioxide (SO_3) from combustion gases at temperatures of at least 350°F by use of pulverized calcium carbonate ($CaCO_3$), Ref. (2). These processes and ideas provided the background for the later large-scale injection of limestone and other sorbents into utility boilers for the purpose of sulfur oxides control.

3.1 General Design and Operation

Reduction of sulfur oxides by direct sorbent injection into a fossil fuel-fired boiler is conceptually a simple process. As depicted in the general flow diagram shown in Figure 1, limestone or other alkaline sorbents are pulverized before injection into the combustion zone of the boiler. Depending on the type of sorbent, particle decomposition, accompanied by evolution of carbon dioxide (CO_2) or water, may occur as the particle is heated, followed by sulfur oxide absorption as the particles travel with flue gas through the boiler. The reacted particles are then removed from the gas stream in dust collection equipment.

Specific details of the sorbent injection point and method have varied in previous tests and trials, and are still under investigation. Important variables to be considered when choosing an injection point and method are the temperature regime at the injection location and the quality of sorbent distribution achieved by the specific injection method. The time and

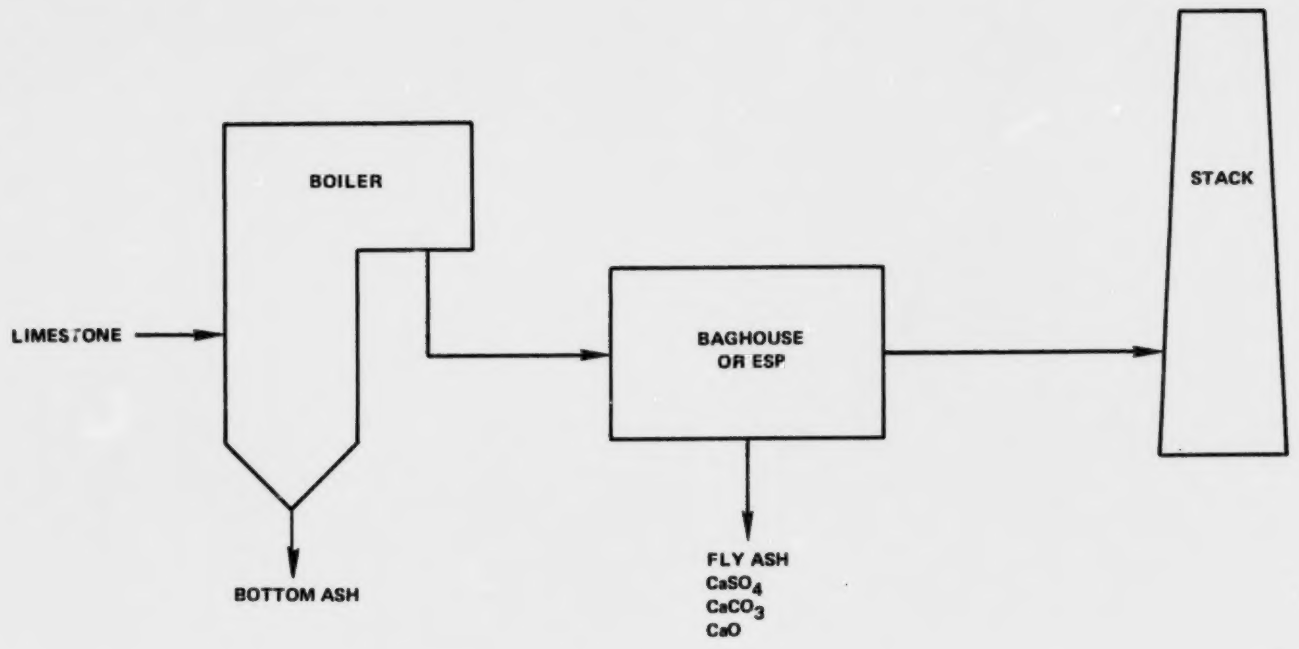

BOILER
LIMESTONE
BOTTOM ASH
BAGHOUSE OR ESP
FLY ASH
$CaSO_4$
$CaCO_3$
CaO
STACK

Figure 1. Flow schematic for LIMB system.

temperature history of a particle and the thermodynamics and chemical kinetics of the process are the primarily determinants of the sorbent reactivity and ability to remove sulfur dioxide from the gas. Possible injection locations are with the fuel, through the air ports, or higher in the boiler.

Because of the difference in hardness between limestone (and most other sorbents) and coal, it is advantageous to mill operation and sorbent fineness control to employ a separate milling circuit for the sorbent. Figure 2 is a flow schematic and equipment description for a typical storage, pulverization, and feeding operation for limestone injection. Material from this system can be transported to any injection location.

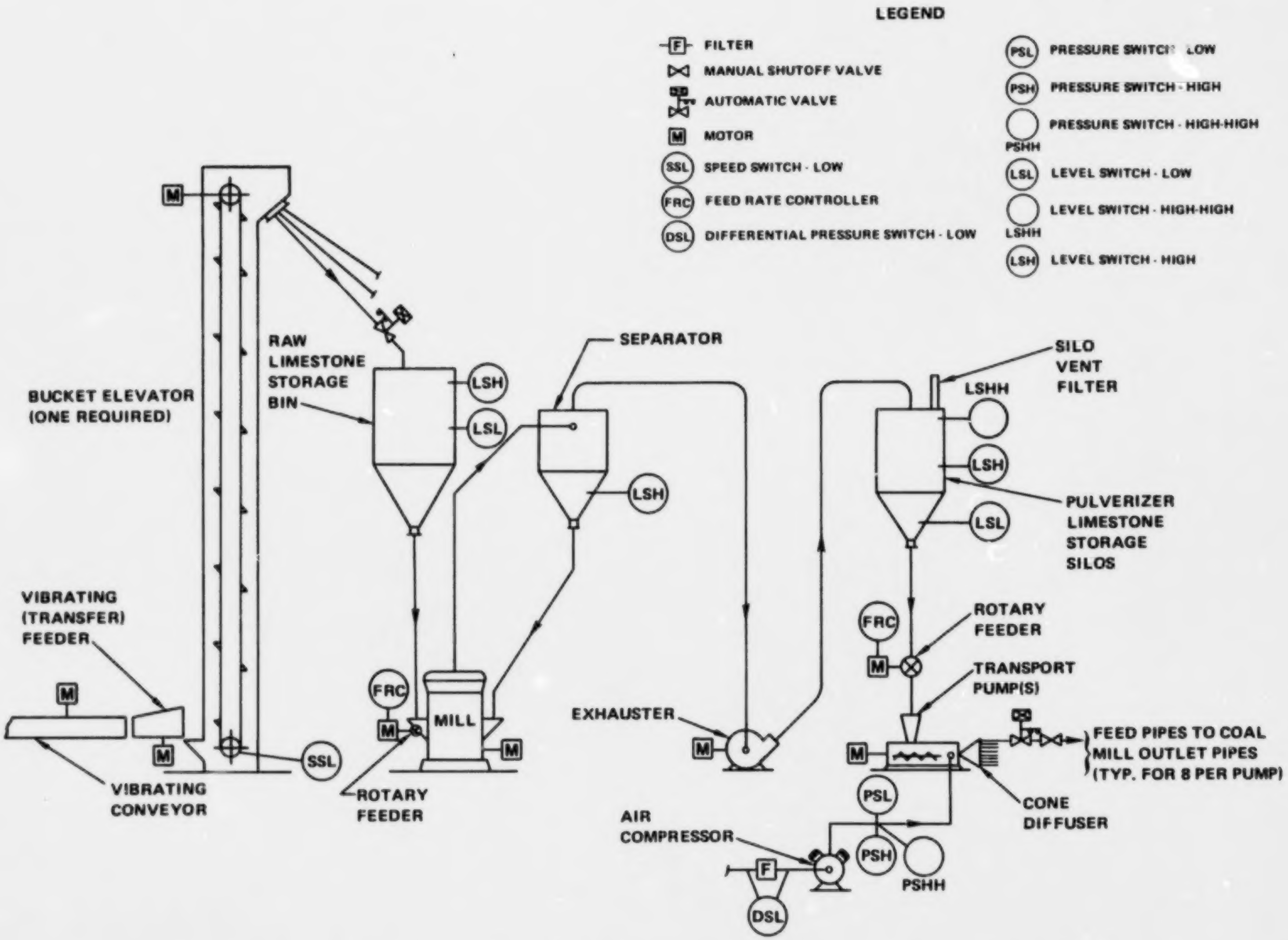
Injection of sorbent into the furnace results in an increase in solids passing through the furnace. The amount of sorbent fed is regulated according to coal sulfur content and emission control requirements. To handle the increased particulate loading, sootblower frequency or number may increase, along with changes in operation of dust collection equipment.

In Section 4, Process Design, additional information pertinent to design and operation of direct injection systems is discussed in the overall review of past and on-going studies of sorbent injection.

3.2 History of Development

The following discussion presents a synopsis of significant work in the area of furnace injection during the past 30 years. This discussion is not intended to be exhaustive, but to generally illustrate the types of testing conducted, the performing organizations, and the past and present objectives and direction of the process.

Noteworthy investigations of furnace injection were begun in the 1950's. One of the first studies was by TVA in a small, coal-burning pilot plant in 1955. The unit was designed primarily to supply gas for an ammonia scrubbing process and did not have the range of temperature and residence time to adequately study limestone injection. Soon after, testing was shifted to


LEGEND
FILTER
MANUAL SHUTOFF VALVE
AUTOMATIC VALVE
MOTOR
SPEED SWITCH - LOW
FEED RATE CONTROLLER
DIFFERENTIAL PRESSURE SWITCH - LOW
PRESSURE SWITCH - LOW
PRESSURE SWITCH - HIGH
PRESSURE SWITCH - HIGH-HIGH
LEVEL SWITCH - LOW
LEVEL SWITCH - HIGH-HIGH
LEVEL SWITCH - HIGH
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M
SSL
FRC
DSL
PSL
PSH
PSHH
LSL
LSHH
LSH
BUCKET ELEVATOR (ONE REQUIRED)
RAW LIMESTONE STORAGE BIN
SEPARATOR
SILO VENT FILTER
PULVERIZER LIMESTONE STORAGE SILOS
VIBRATING (TRANSFER) FEEDER
VIBRATING CONVEYOR
ROTARY FEEDER
MILL
EXHAUSTER
ROTARY FEEDER
TRANSPORT PUMP(S)
FEED PIPES TO COAL MILL OUTLET PIPES (TYP. FOR 8 PER PUMP)
CONE DIFFUSER
AIR COMPRESSOR

Figure 2. Limestone preparation and injection system (Ref. 30).

small-scale tests with simulated stack gas or air-sulfur dioxide mixtures. Limestone and dolomite were used as sorbents. Limestone was found to be more effective than dolomite and produced removal efficiencies of 20% to 70% as the mole ratio of calcium to sulfur varied from 1:1 to 4:1.

In the early 1960's, considerable development work was performed by Wickert of West Germany, as well as Wahnschaffe and associates, Ref. (3), of Volkswagenwerk in Wolfsburg, West Germany. At Volkswagenwerk, a hydrated, calcined dolomite was injected into a boiler just downstream of the combustion zone. Considerable success was achieved in the removal of SO_2. No boiler fouling problems were encountered. No appreciable amount of SO_2 removal was measured when limestone was used as the reactant. These tests were later discontinued due to the high cost of raw material in the plant area.

During the latter part of the 1960's, further investigations were conducted in Japan, Europe, and the United States. Both pilot- and full-scale investigations were conducted by the Central Research Institute of Electric Power Industry in Japan, under the direction of Y. Ishihara, Ref. (2). Limestone performed well in the pilot plant, producing 32% to 40% sulfur removal with a stoichiometric calcium to sulfur feed rate. In the plant tests, hydrated lime was found to perform better than limestone.

In Essen, West Germany, Bergbau-Forschung GmbH, Ref. (4), constructed a 23-foot by 4-inch burner tube to study limestone and dolomite injection. At temperatures up to 1650°F, 80% to 90% SO_2 removal was obtained for stoichiometric ratios of 3:1.

In the United States, bench-scale research was conducted by Battelle Memorial Institute, Ref. (5), under contract to the National Air Pollution Control Administration. In this study, primary emphasis was on sorbent properties and their behavior under various time/temperature conditions, both with respect to calcination and SO_2 absorption capabilities.

Under sponsorship of the National Center for Air Pollution Control and later the Environmental Protection Agency, R. E. Borgwardt and associates, Ref. (6,7), conducted bench-scale tests on the reaction kinetics of various limestones and dolomites during the late 1960's and early 1970's.

In larger scale tests during this time period, Wisconsin Electric Power Company in 1964 initiated studies of direct injection at their Port Washington, Wisconsin plant. Dolomitic limestone was mixed and fed with the coal. Most tests were conducted at about 130% of the stoichiometric calcium to sulfur ratio. Removal averaged 50% at this feed rate for coal with 2.8% sulfur. Deposits were formed on the superheater tubes which could not be removed by sootblowing.

Also on a commercial scale in 1966, Combustion Engineering, Ref. (8), in conjunction with Detroit Edison, conducted field tests on the injection of both dolomite and limestone into one furnace of a 325 MWe twin furnace unit, followed by wet scrubbing. While introducing sufficient sorbent to react with all the sulfur in a 2.5% to 3.5% sulfur coal, in-furnace removal efficiencies of about 25% were obtained. No adverse effects on furnace operation were recorded.

Following the C-E tests, the Tennessee Valley Authority, under contract to the U.S. Environmental Protection Agency, conducted limestone injection tests on a 150 MWe pulverized coal-fired boiler at TVA's Shawnee Steam Plant, Ref. (9). In contrast to the earlier bench-scale results and the preliminary findings at Detroit Edison, overall SO_2 removal performance was poor and furnace fouling was significant.

Further work conducted by Combustion Engineering at Kansas City Power and Light, Ref. (10), Kansas Power & Light, Ref. (11,12), and Union Electric, Ref. (12,13), with furnace injection followed by wet scrubbing, was more in line with the TVA results. Although furnace fouling was not a problem in either case, being dependent on coal ash properties and boiler design, sorbent reactivity was a problem at both locations and appeared to be reduced by "deadburning" of the sorbent, caused by calcination at excessive temperatures. Furnace injection was subsequently abandoned at both power plants.

In 1970, Babcock and Wilcox, Ref. (14), issued a research report covering extensive pilot plant work done under National Air Pollution Administration funding. In the pilot study, a total of 415 tests were conducted with 129 different sorbents and seven pulverized coals. Up to 60% removal was obtained at stoichiometric ratios ranging up to 4:1. Many process variables were investigated. Based on the behavior of the ash-sorbent particulate in these tests, it was predicted that furnace fouling would not be a problem. However, resistivity measurements of the particulate matter emitted indicated that electrostatic precipitator performance could be adversely affected.

During the remainder of the 1970's, interest in direct sorbent injection subsided and was replaced by development of other flue gas desulfurization techniques. However, investigations have again begun, initiated by changes in boiler operating conditions which have the effect of lowering combustion zone temperatures. The changes are due to the high moisture content of low rank coals currently being used and the fact that recently designed boilers are larger to accomplish the same heat transfer. This combination of lower combustion temperatures and increased residence time of the combustion gases in the boiler has proven to be beneficial for SO_2 removal by direct sorbent injection. Hein and associates, Ref. (15,16), at the Rheinisch Westfalishes Elektrizitatswerk, Essen, West Germany, have reported test results from sorbent injection in a 60 MWe boiler. At a stoichiometric ratio of 2:1, better than 75% SO_2 removal was accomplished, and no boiler slagging or fouling problems were reported.

Because of regulations limiting NO_x emissions, work was also performed in recent years to develop combustion techniques for reducing the formation of NO_x. The techniques involve controlling the amount and location of primary and secondary combustion air and have the overall effect of reducing peak flame temperatures. While operating these burner systems, investigators found that the reduced flame temperature and fuel-rich combustion conditions improve SO_2 removal by direct sorbent injection.

In tests conducted by Energy and Environmental Research for the EPA, Ref. (17), results from low-NO_x burner facilities have demonstrated the importance of the temperature history experienced by the sorbent. Sulfur capture was affected by design parameters, sorbent injection location, and thermal environment, and was facility dependent.

Mitsubishi Heavy Industries, Ref. (18), under contract to the Electric Power Research Institute (EPRI), recently completed a study of limestone furnace injection using a low-NO_x burner. In preliminary reactor tube tests, it was concluded that hydrogen sulfide (H_2S) absorption from fuel-rich conditions was not as effective as SO_2 absorption from a gas containing excess oxygen (O_2). In the 4-ton/hour test furnace, 30% to 40% SO_2 removal was obtained with high sulfur coal (3.1%) and low sulfur coal (0.75%) at a limestone stoichiometry of 2:1 under low-NO_x firing conditions.

Combustion Engineering, Inc, Ref. (19), recently completed a 4×10^6 Btu/hr pilot scale test for EPRI in which furnace slagging and fouling behavior with limestone injection during low-NO_x firing conditions was investigated. Two coals were evaluated. For the case of a high sulfur Eastern coal, slagging behavior and deposit friability improved with limestone addition, although deposit buildup was at a faster rate. Slagging and fouling deposit characteristics also benefited from limestone addition for a high sodium Western subbituminous coal. In both cases, SO_2 removal efficiency was dependent on particle residence time at favorable temperatures.

In recent tests reported by Radian Corporation, Ref. (20), limestone was injected into a 52 MWe tangential firing lignite boiler at Hoot Lake Station of Otter Tail Power Company. Most tests were conducted in a low-NO_x firing mode. Sulfur dioxide removal ranged from 20% to 50% at stoichiometries from 1.6:1 to 5:1. Boiler fouling behavior was improved.

SECTION 4

PROCESS DESIGN

This section provides a more detailed description of the physical and chemical factors which influence the capture of SO_2 by limestone and other sorbents. These factors are divided into three categories: (1) those which are properties of the coal; (2) those which are properties of the sorbent; and, (3) those which are related to the physical and chemical properties of the furnace environment. A discussion of those factors, plus a description of necessary considerations for the sorbent preparation system, is presented in the following subsections.

4.1 Coal Properties

The design and operation of a pulverized coal-fired steam generator depends heavily on the physical properties of the selected coal. The mineral matter composition can vary significantly among the various coal types. The anticipated behavior of this mineral matter during operation has a major impact in defining the configuration of the boiler. Sorbent addition will alter the behavior of the mineral matter with respect to the rate of deposit and ease of removal. The major coal properties are discussed in the following subsections. Additional information pertaining to sorbent/coal interaction can be found in Section 5.

4.1.1 Types

There are five basic types of coal which are presently being used in the United States and which will continue to be used for at least the next 25 years. These coals are Eastern and Midwestern bituminous (high and low sulfur), subbituminous, Texas lignite, and North Dakota lignite (low sulfur). Dry sorbent injection will impact each of these coals differently, generally as a function of sulfur content and coal ash properties.

Coal contains a sizeable percentage of noncombustable ash material, the amount varying with the coal grade. When fired, this material can accumulate on the boiler tubes and, if uncontrolled, can reduce heat transfer and, consequently, boiler efficiency. The ash properties are important in determining the performance and slagging and fouling tendencies of a particular coal. The most important ash properties are generally considered to be:

Ash fusibility temperature
Base/acid ratio
Silica/alumina ratio
Iron/calcium ratio
Active alkali content
Sodium content
Gravity fractionation

4.1.2 Ash Fusibility Temperature

Fusibility temperature is an actual measurement under laboratory controlled conditions showing the melting behavior of coal ash. It is perhaps the most basic means for predicting ash performance under furnace conditions. As measured by ASTM Standards, four points are measured:

1. Initial deformation temperature (IT)
2. Softening temperature (ST)
3. Hemispherical temperature (HT)
4. Fluid temperature (FT)

Both the temperature levels and temperature spread between IT and FT are important in predicting ash behavior. For example, a lowered ash ST indicates a greater tendency for ash deposits buildup, whereas a wider IT-FT spread usually indicates greater ease of removal of the bulk of the deposit. As the IT-FT spread narrows, the removal of deposits may become more difficult.

4.1.3 Base/Acid Ratio

The base/acid ratio (B/A) has been empirically correlated with ash melting temperatures. This index is defined as the ratio of the ash base salt content to the ash acid salt content. Base salts are ferric oxide (Fe_2O_3), lime (CaO), magnesium oxide (MgO), sodium dioxide (Na_2O), and potassium oxide (K_2O). Acid salts are silica (SiO_2), alumina (Al_2O_3), and titanium dioxide (TiO_2). Thus,

$$B/A = \frac{Fe_2O_3 + CaO + MgO + Na_2O + K_2O}{SiO_2 + Al_2O_3 + TiO_2}$$

While the ash acid salt constituents are usually considered to be high melting, ash fusion temperatures are lowered proportionally by the relative amounts of basic oxides available in the ash for reaction. When used in conjunction with ash fusibility temperature and other data, base/acid ratio can be valuable in predicting the relative performance of coal ash in a furnace. Minimum melting points and higher slagging potential are generally associated with B/A's of 0.4 to 0.7.

4.1.4 Silica/Alumina Ratio

The silica/alumina ratio ($S/A = SiO_2/Al_2O_3$) provides additional information relating to ash fusibility. Although both of these constituents are acidic, SiO_2 is more likely to form lower melting species with the basic constituents than is Al_2O_3. Thus, a higher S/A ratio indicates a lower fusibility temperature.

According to data obtained at C-E's Kreisinger Development Laboratory, under reducing atmospheres (less than stoichiometric), no appreciable change in ash fusibility is noted for S/A ratios in the range of 1.7 to 2.8. For decreasing ratios less than 1.7, the ash softening and fluid temperatures both increase; for ratios greater than 2.8, the fluid temperature decreases.

4.1.5 Iron/Calcium Ratio

A single ratio such as base/acid cannot completely define the characteristic behavior of a coal ash. Base/acid ratio does not, for example, account for differences in fluxing action among the several basic components. (Fluxing refers to the property of the component or substance which lowers the fusion temperature of the mixture.)

Among the five basic-oxide ash components, iron and calcium are the most important, primarily because they make up the largest portion of the basic constituents. The combined fluxing effect of iron and calcium oxides has a complex relationship to the percentage of the two fluxes present. In general, this effect is most pronounced when the iron/calcium ratio ($I/C = Fe_2O_3/CaO$) is between 3 and 0.3.

4.1.6 Active Alkali and Sodium Content

Alkali metals play important roles in ash deposition. Besides vaporizing and condensing to form deposits, they act as strong fluxing agents, reducing melting temperatures. Potential fouling does not correlate well with total ash alkali content. Rather, the nature of the alkali is the influencing factor. The total alkalies can be separated into two general categories:

- Active alkalies that readily vaporize, react, and condense. These alkalies can be extracted from coal by leaching in a weak solution of acetic acid.

- Inactive alkalies which are more stable and remain relatively inert. These alkalies are not extracted by leaching.

Sodium, an active alkali, has been found by C-E to correlate well with convective section fouling. The ash deposit bonding strength usually increases as sodium content increases. The measure of sodium content is not the total sodium in the ash but the soluble sodium, which may be substantially different.

Western low rank coals, which generally have high alkali ash, correlate particularly well with the amount of active sodium in the ash:

% Soluble Na_2O in Ash (Equivalent Soluble Sodium in Ash)	Fouling Tendency
1.0	Low
1.0 - 2.0	Medium
3.0	High
5.0	Severe

With Eastern coals, which are characterized as having low alkali content, the chloride content, because of its relative ease of determination, is frequently used as an indicator of the fouling potential of the active alkalies. American experience confirms that tightly bonded deposits in the convective passes can be expected when the inorganic chloride level in the coal exceeds 0.3 percent.

4.1.7 Gravity Fractionation

Because segregation of mineral matter is a function of particle size, it is important to analyze coal in the size range at which it will be used in predicting behavioral tendencies of the mineral matter during combustion. Gravity fractionation has been developed for this purpose.

A sample of pulverized coal is separated by density, using various organic liquids having specific gravities ranging from 1.3 to 2.9. After air drying, the samples are analyzed for percent ash, ash composition, and ash fusibility. One of the more significant compositional determinations is the distribution of the various iron compounds.

4.1.8 Slagging and Fouling

Slagging is the deposition of molten or partially fused ash particles on tube surfaces. The particles are carried by the gas to the furnace walls as well as up to the radiant surfaces in the upper furnace. When the particles strike a cool tube surface, they chill and solidify. Radiant surfaces in the furnace are most susceptable to slagging, but convective surfaces can also slag if gas temperatures exceed prudent levels. Normally, coals with low ash fusion temperatures have a high potential for slagging.

Fouling results from the condensation of the volatile coal ash constituents which have been vaporized during combustion (e.g., sodium sulfate) onto flyash particles and boiler tubes in areas where temperatures are below the dew points of the constituents.

4.2 Sorbent Properties

Several of the sorbent properties which influence the effectiveness of SO_2 capture and utilization of the sorbent are sorbent type (i.e., lime, limestone, sodium carbonate, etc.), calcination reactions (if applicable), particle size, porosity, and time/temperature history experienced by the sorbent. The sorbent also influences overall SO_2 capture according to the calcium to sulfur stoichiometry employed and the quality of dispersal of the sorbent in the flue gas.

4.2.1 Types

Sorbents used for furnace injection are usually calcium, magnesium, or sodium alkalis. By far the most common alkalis employed have been calcium-based, and of these compounds, calcium carbonate (limestone) has been the most extensively used. A review of available literature on the relative sulfur removal efficiency of several types of the more common calcium compounds produces the following consensus ranking, in decreasing order:

1. Hydrated lime ($Ca(OH)_2$)
2. Dolomitic limestone ($CaCO_3 . MgCO_3$)
3. Limestone ($CaCO_3$)
4. Lime/calcined limestone (CaO)
5. Marl
6. Shell
7. Trona

With the exception of lime and hydrated lime, these products occur naturally and are obtained by mining. Limestone is by far the most widely available product occurring in virtually every state, although not always in accessible locations.

The general term "limestone" covers a broad range of materials varying in calcium and magnesium content--from calcite or aragonite ($CaCO_3$) to dolomite ($MgCO_3 . CaCO_3$). The terms "high calcium limestone" (very low in magnesium carbonite ($MgCO_3$) and "dolomitic limestone" (approaching dolomite in $MgCO_3$ content) are commonly employed. $MgCO_3$ content affects SO_2 absorption because it does not behave like $CaCO_3$, either in regard to calcination or reaction with SO_2.

Limestones vary widely in their chemical composition and crystal structure, with the result that material from one source may differ from another in degree of porosity developed during calcination and, therefore, in absorption efficiency. High quality, high calcium limestones contain 90% to 99% $CaCO_3$, with SiO_2 and Al_2O_3 comprising the major impurities. Dolomitic limestone typically contains about 40% to 43% $MgCO_3$, 1% to 3% impurities, and the rest $CaCO_3$. The bulk density of limestone varies between 125 to 175 lb/ft^3, with dolomitic limestone being slightly higher.

Shell is a high purity fossiliferous calcite of marine origin, primarily oyster and clam shells. There may be some small amounts of impurities, principally silica sand. This material is limited to a few coastal areas.

Marl is also of marine origin, with varying amounts of sand and clay impurities mixed with calcium deposits. It is older and less pure than shell.

Trona ($NaHCO_3 \cdot Na_2CO_3 \cdot 2H_2O$) or sodium sesquicarbonate, is primarily suited for post-furnace dry absorption. Injected at the inlet to a fabric filter, it would collect on the bags. The sulfur dioxide (SO_2) would react with the Trona as it passed through the baghouse. The resulting sulfate material would be removed with the flyash. This naturally occurring material is limited in availability to small sections of the Southwest.

Lime is obtained by driving off the carbon dioxide (CO_2) from limestone in a kiln (calcining). The resulting powder is relatively stable, although it is very hygroscopic. By hydrating lime with enough water to satisfy its chemical affinity, $Ca(OH)_2$ is formed. This powder has a strong affinity for CO_2. When subjected to temperatures of 1000°F, $Ca(OH)_2$ will decompose to CaO and H_2O.

Hydrated lime is generally considered to be the most effective calcium-based sorbent for in-furnace injection. However, limestone could have advantages as a sorbent for several reasons. First, limestone will not require pre-calcination, which increases the initial cost. Second, although calcination of the raw limestone will require extra heat input to the boiler (600 Btu/lb limestone), this is more than offset by the higher cost of hydrated lime for the same sulfur removal efficiency. Finally, calcined limestone has a tendency to deadburn in the boiler, resulting in an increase in the amount of sorbent required to achieve a given sulfur removal.

4.2.2 Time/Temperature Effects on Calcination and Porosity

For the three compound types mentioned--calcium, magnesium, and sodium--the calcination reactions are as follows:

$$CaCO_3 \xrightarrow{Heat} CaO + CO_2$$

$$MgCO_3 \xrightarrow{Heat} MgO + CO_2$$

$$Na_2CO_3 \xrightarrow{Heat} Na_2O + CO_2$$

In an early fundamental study by Battelle, Ref. (5), the equilibrium dissociation pressure of carbon dioxide (CO_2) maintained by limestone was plotted as a function of temperature. These data, shown in Figure 3, demonstrate that a temperature of about 1400°F is necessary to keep the calcined limestone from recombining with CO_2 at a typical furnace CO_2 concentration of 14.5%. Based on thermodynamics, it was also concluded by Battelle that dolomite dissociates at 865°F according to the reaction:

$$CaCO_3.MgCO_3 \xrightarrow{Heat} CaCO_3 + MgO + CO_2$$

However, similarities are reported to exist between decomposition of limestone and dolomite. CO_2 is stated to retard decomposition of both and H_2O to accelerate it. Impurities can affect decomposition of both and also the rate of deadburning of the oxide.

Following the fundamental study by Battelle, pilot-scale work was conducted in the reactor shown in Figure 4. The reactor was one foot in diameter and five feet long, with a gas velocity of 5 fps. The effects of residence time and temperature on calcination of limestone are shown in Table 1. The data indicate that the calcination process was not carried to completion, except possibly in one run at the highest temperature. The lack of complete calcination might be due either to the failure of the particles to attain the listed reaction temperatures or to the inherent kinetics of calcination. However, it is likely that the kinetics of calcination is the limiting factor since particle heat-up is predicted to be very rapid. The calculated time for a 90 micron limestone particle to attain a temperature close to the gas phase is shown in Figure 5 to be on the order of 0.05 seconds.

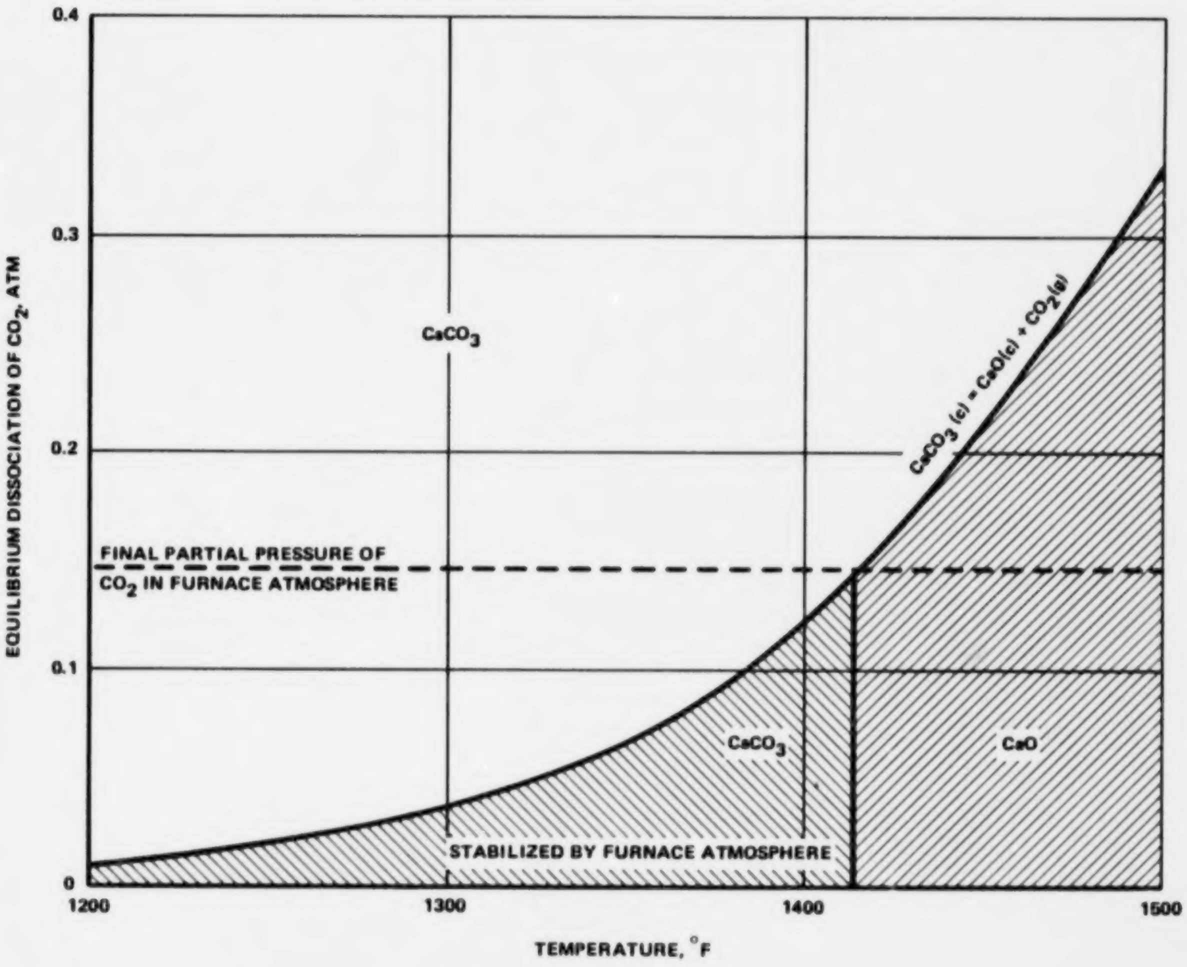


Figure 3. Effect of furnace atmosphere on dissociation of $CaCO_3$ (Ref. 5).

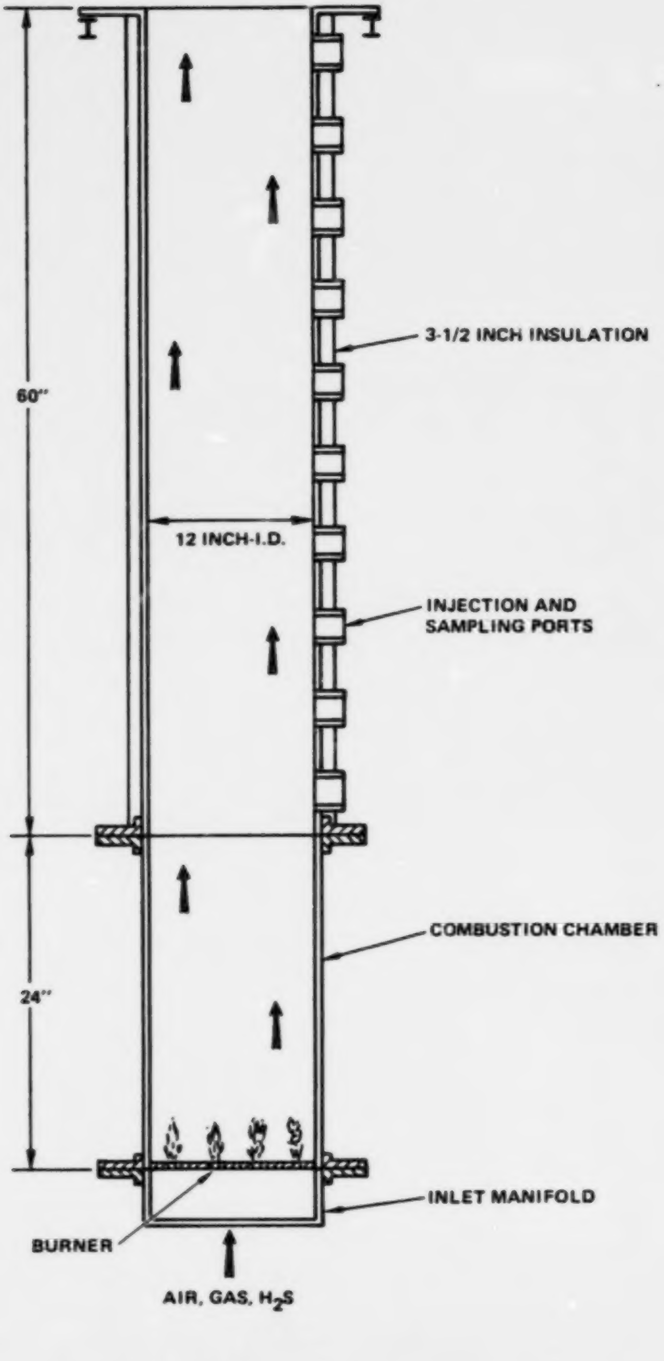

60"
12 INCH-I.D.
3-1/2 INCH INSULATION
INJECTION AND SAMPLING PORTS
24"
COMBUSTION CHAMBER
INLET MANIFOLD
BURNER
AIR, GAS, H_2S

Figure 4. Reactor Configuration (Ref. 5).

TABLE 1. CALCINATION OF LIMESTONE (REF. 5)

	Conditions in Reactor						
Run Number	Reaction Temp, °F	Residence Time, sec	Weight Percent CO_2	Weight Percent $CaSO_4 \cdot 2H_2O$	Percent CaO Utilization	Percent Calcination	Percent Reaction of CaO Formed
5118	1595	0.31	41.4	2	1.1	7.9	14
5010	1595	1.01	31.8	7.5	3.9	35.9	11
4006	1701	0.27	38.5	4	2.2	16.7	13
4007	1701	0.27	38.5	4	2.2	16.7	13
4005	1701	0.85	18.7	14	6.3	66.9	9
3304	1993	0.15	32.6	8	4.2	33	13
3606	2000	0.56	2.5	32	13.7	96	14

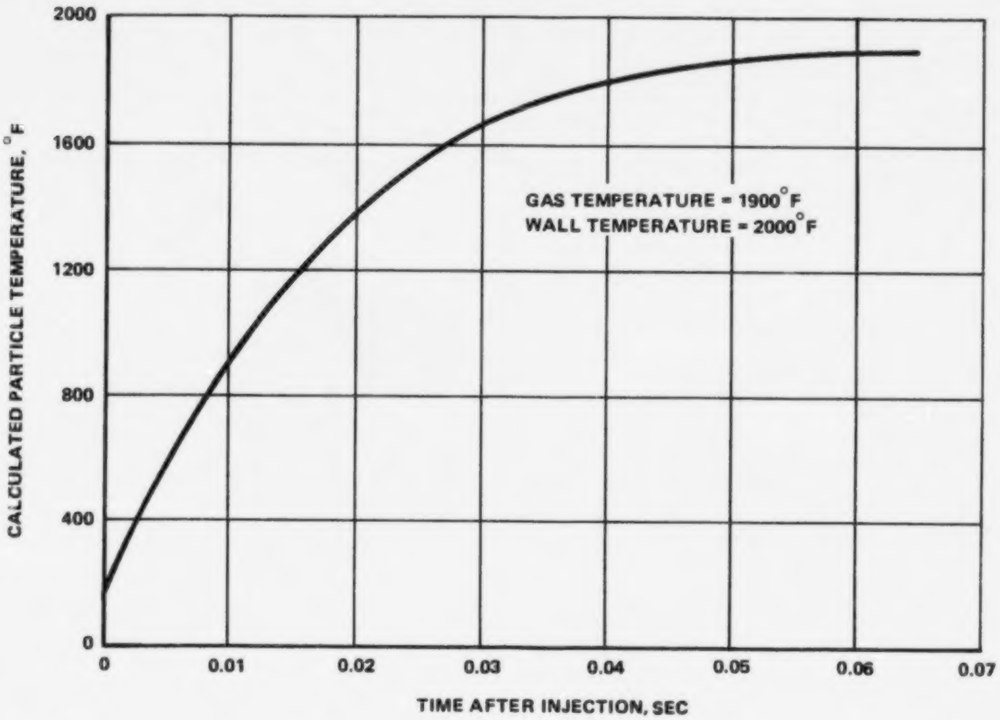


Figure 5. Heat-up time of a 90-micron limestone sphere (Ref. 5).

In addition to the kinetics of sorbent calcination, temperature and residence time play a major role in determining final particle size, porosity, and, therefore, total available surface area. Based on reports of Battelle and Borgwardt, Ref. (5,7), the most important factors in determining the reactivity of a given limestone are the surface area and porosity developed during calcination. These factors directly determine the availability of lime for reaction, the accessibility of the lime, and the space which is available within a given particle for reaction products. The surface area and porosity are determined by calcination rate and sintering rate. The latter is more important in determining the decreased reactivity of limestones at high temperatures. Sintering itself is a complex process, depending on grain size, impurity content, lime temperature and calcination time.

Data shown in Table 2 summarize surface area and porosity measurements by Battelle on limestones calcined under a variety of temperature and residence time conditions. The data indicate that two magnitudes of pore systems exist, the smaller with diameters on the order of a few tenths of a micron. This conclusion was reached due to the samples having larger pore volumes than expected, and was thought to be caused by cracks in the lime from the violent nature of the calcination process. Maximum surface areas appeared to occur with injection temperatures in the range of 1900°F to 2000°F, implying more overburning at higher temperatures.

In order to clarify the effects of temperature on surface area development, additional tests were run at similar residence times. The results, given in Table 3, show that surface area decreases with increasing temperature.

In a paper recently presented by Borgwardt, Ref. (21), the surface area of calcined limestone was determined as a function of initial particle size in one series of tests and as a function of calcination time in another. These results are shown in Tables 4 and 5. The significant conclusion from both tests is that the final surface area is uniform regardless of initial particle size.

TABLE 2. POROSITY AND SURFACE AREA DATA (REF. 26)

Sample[a] Number	Temperature, °F		Residence Time, sec	Percentage Calcination	Pore Volume[b], cc/g CaO	Surface Area, m^2/g CaO	Average Pore Diameter, microns
	Injection	Collection					
925-B	2224	1440	2.86	86	0.187	21.3	0.044
925-A	2224	1765	1.27	83	0.203	21.5	0.048
925-D	1830	1440	1.82	73	0.337	61.3	0.028
925-C	1830	1765	0.23	19	0.575	44.2	0.066
106-A	2243	2156	0.15	40	0.133	52.9	0.013
106-B	2156	2073	0.16	45	0.226	66.7	0.017
106-C	1996	1920	0.17	28	0.411	67.8	0.030
106-F	1920	1500	1.35	78	0.164	36.6	0.022
106-E	1849	1782	0.18	45	0.166	42.9	0.019
106-D	2243	1500	0.99	85	0.100	7.6	0.066

[a] Series 925 is -140/+200 mesh; series 106 is -270/+325 mesh

[b] For pores with $d < 2\mu$

TABLE 3. EFFECT OF TEMPERATURE ON SURFACE AREA (REF. 26)

Run	Stone[a] Type	Temperature, °F		Residence Time, sec	Percentage Calcination	Area[b], m^2/g
		Injection	Collection			
318-A	Mich. Marl	2006	1712	1.03	52	13.6
318-B	IGS-4	2006	1712	1.03	72	24.8
318-C	No. 1336	2006	1712	1.03	71	26.4
317-A	Mich. Marl	2155	1785	1.17	66	7.5
317-B	IGS-4	2155	1785	1.17	80	16.3
317-C	No. 1336	2155	1785	1.17	83	23.7
318-D	Mich. Marl	2263	1828	1.14	74	5.9
318-E	IGS-4	2263	1828	1.14	81	8.2
318-F	No. 1336	2263	1828	1.14	84	14.1

[a] -140/+200 mesh

[b] Area per gram of sample

TABLE 4. BET SURFACE AREAS-LIMESTONE CALCINED IN RTP FLOW REACTOR (REF. 21)

Reactor Temperature, °C	Particle Size, μ	Calcination, %	Surface Area, m^2/g
925	10	78	51.5
1000	10	85	46.8
1000	10	88	49.7
1000	50	82	45.7
1075	10	90	45.2
1075	90	75	38.2
1075	90	84	41.3

TABLE 5. BET SURFACE AREAS-1.6μ LIMESTONE CALCINED IN RTP DIFFERENTIAL REACTOR (REF. 21)

Reactor Temperature, °C	Calcination Time, min	Surface Area, m^2/g
650	23.00	49
700	7.50	45
750	2.50	40
800	1.20	36
850	0.92	40

In contrast to the surface areas shown in the tables of Borgwardt (40 to 50 m^2/gm), surface areas of kiln-calcined limes are generally on the order of a few square meters per gram or less. Borgwardt attributes the difference in surface areas to the difference in calcination environment between kiln and dispersed systems. With respect to rate of calcination, he concludes it is determined by the specific BET surface area of the solid.

In other studies, Hein et al, Ref. (15), reported significant decomposition and surface area results depending on the original composition of the alkali. The best temperature for calcination, in terms of the amount of particle surface area produced, varied according to sorbent type, as shown in Figure 6. Specific surface areas peaked at 1242°F, 1832°F, and 1832°F (700°C, 1000°C, and 1000°C) for hydrated lime ($Ca(OH)_2$), $CaCO_3$, and $CaCO_3 \cdot MgCO_3$ (dolomite), respectively. The decay in specific surface area experienced for all three sorbents after having reached peak value indicates the effect of dead burning or sintering, which is time and temperature dependent.

Also, dolomitic stones were found by Hein to perform better overall than pure $CaCO_3$. It is postulated that this increase in performance is due to $MgCO_3$ decomposing at lower temperatures than $CaCO_3$, therefore contributing to particle breakup. With respect to the various calcium alkalis, Klaus and Glaser, Ref. (22), found that performance, with decreasing efficiency, was: $Ca(OH)_2$, $CaCO_3$, and CaO. The top performance of $Ca(OH)_2$ is believed to be due to its low temperature decomposition, causing particle breakup and porosity changes. CaO performed worse because no decomposition occurs in the furnace, thus allowing surface sintering or "deadburning".

Since the $MgCO_3$ in dolomite calcines at lower temperatures, its beneficial effect on particle breakup is possibly at the expense of deadburning the magnesium oxide (MgO) formed. However, as will be discussed in a later section, due to the fact that magnesium sulfate ($MgSO_4$) decomposes at a lower temperature than calcium sulfate ($CaSO_4$), its effectiveness as an SO_2 absorbent is also limited for thermodynamic reasons.

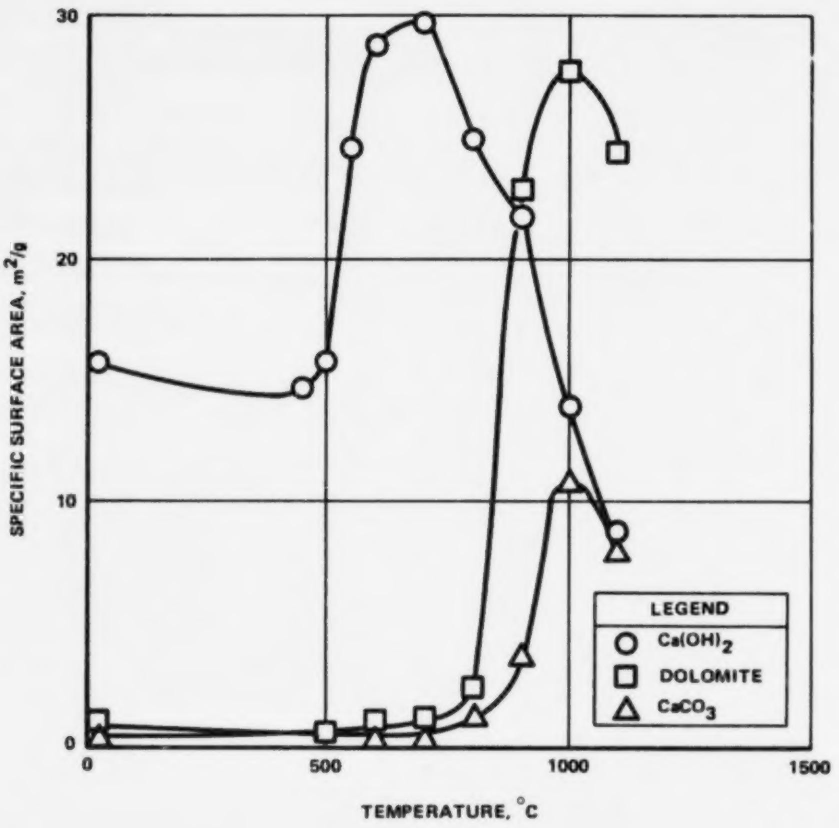


Figure 6. Specific surface area (m^2/g) as function of temperature (laboratory tests, 1-minute sample) (Ref. 15).

4.2.3 Particle Size

In addition to porosity changes caused by calcination, the initial particle size of the sorbent also has an effect on the final available surface area. The importance of this effect, however, was different in the various studies. For example, results reported by Borgwardt, Ref. (21), show that final surface area was unaffected by initial particle size in the 10μ to 90μ size range, but results of SO_2 absorption studies have generally shown increasing performance with decreasing particle size, at least down to specific lower limits.

In the Battelle studies, Ref. (5), microscopic examination of reacted lime particles and electron-microprobe scans made on the same materials showed that the reaction between lime and SO_2 is confined largely to the external surface of the particles. This was indicated also by the fact that the interior of reacted particles were soluble in acid, whereas the outer portion was insoluble. However, because of the pore structure, the interior of the particles was not totally excluded from reaction. It was therefore concluded that the reaction rate and the capacity for reaction depend both upon particle size and pore structure.

In specific tests conducted, stone samples were prepared by classifying with U.S. Standard sieves, and contained a distribution of particle sizes and shapes. In general, the results indicated an increase in rate of reaction with decreasing particle size. A plot of rate of reaction versus the average size of the mesh opening for the stone samples is reproduced in Figure 7. In interpreting these results, Battelle concluded that although the rate was increased, the overall utilization of smaller particles did not increase significantly.

Other investigators have reported increases in overall absorption efficiency with smaller particles, indicating an increase in particle utilization as well. Zentgraf, referenced by TVA, found in plant tests that one-third more 70%, -90μ material was required to get the same absorption as

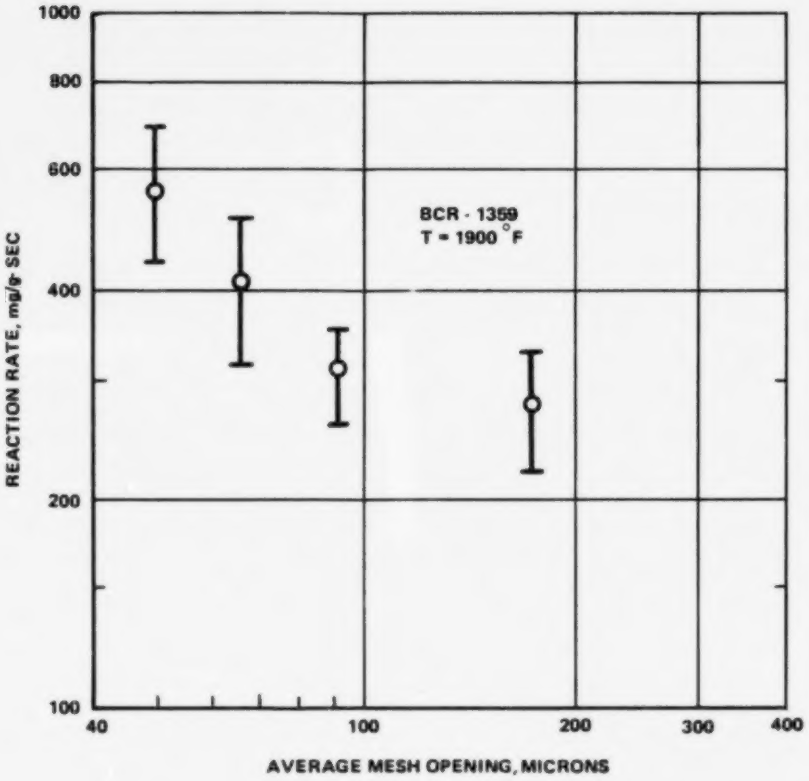


Figure 7. Effect of particle size on reaction rate in reactor (Ref. 5).

with 95%, -90μ material. Although Zentgraf found that absorption was better with smaller particles, it did not increase proportional to surface area. This was attributed in part to agglomeration. Performance leveled off below 50μ to 60μ.

In the B & W studies, Ref. (14), the specific surface of sorbents was measured using the Lea-Nurse technique. The data indicated that maximum reactivity was reached when the surface area was about 2000 cm^2/gm. This surface area corresponded to the -325 mesh (44 μ) size fraction. This size fraction would have an average particle size of 10μ to 15μ.

During the large-scale TVA tests, Ref. (9), a "fine" and "coarse" limestone were used. Fine had 97%, -400 mesh (38 μ), and coarse had 58%, -400 mesh. At high stoichiometries (3:1), "fine" was better than "coarse", but not much difference was observed at lower stoichiometries.

In the earlier TVA report, Ref. (27), the data of Ishihara is reviewed with the following information on particle size. Removal decreased inversely with the one-fourth power of particle size, the results of which are shown in Figure 8. Also contained in the figure is information on Fe_2O_3 content, which was noted to produce increased SO_2 removal efficiencies when present in concentrations over 1.0%. This phenomenon will be discussed further in Section 4.3.4 in which the effects of oxidation catalysts on SO_2 absorption are described.

In summary, sorbent calcination and reactivity as related to particle size and porosity is a complex phenomenon. Most investigators have observed a leveling off in performance as particle size is decreased. This may be due to the interaction of pore diameter and particle diameter. For larger particles, pore diameter may be important while for smaller particles, the effect is minimized. Another factor which may limit the effectiveness of smaller particles is mass transfer. The fine particles should have less slippage in the gas stream than larger ones, and less relative motion between solid and gas. Zentgraf postulated this mechanism as a possible factor in his observations on the effect of particle size.

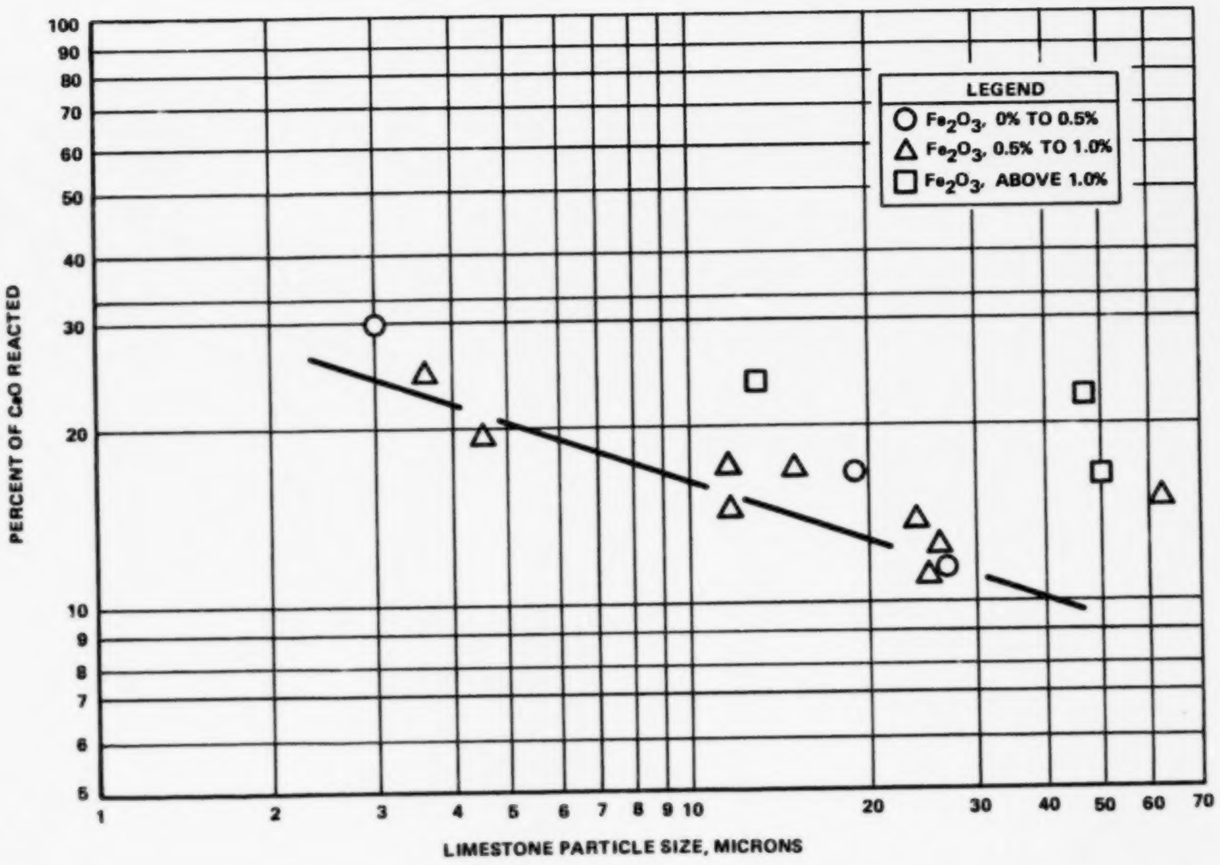


Figure 8. Relation between rate of reacted lime and particle size of limestone (Ref. 27).

4.2.4 Stoichiometry

Sorbent stoichiometry, the amount of sorbent injected per unit of sulfur in the gas (both on a molar basis), influences the total amount of SO_2 absorbed in the furnace injection process and can have a significant effect on the economic viability of LIMB for a particular unit. In a recent presentation by EER, Ref. (17), results from past investigations on pilot- and full-scale systems were plotted together (Figure 9) to demonstrate the effects of sorbent stoichiometry on SO_2 removal. High- and low-sulfur coals can require substantially different amounts of sorbent per mole of sulfur. The selected ratio determines the impact on all other components and subsystems, including gas cleanup, sorbent preparation, and the boiler itself.

Sorbent injection rates at very high stoichiometries (up to 7:1) were tested very early (1955) in a TVA study and are shown in Figure 10. In later tests at TVA's Shawnee Steam Plant (which the data point in the EER plot represents), results were somewhat decreased compared to the earlier study. In these tests, limestone was injected into a 150 MWe pulverized coal fired boiler, the coal having an average sulfur content of 2.7%. At a stoichiometric ratio of 2:1, SO_2 removal ranged from 17% to 20% for the various limestones injected at full-load operation. Removal efficiency did not exceed 40%, even at stoichiometric ratios up to 5:1. In experiments by Wickert in 1963, referenced by Kruel and Juntgen in Ref. (4), on an oil-fired boiler, injection of calcined dolomite at a stoichiometric ratio of 1:1 resulted in desulfurization efficiencies of 15% to 45%. At tests on another oil-fired boiler in Wolfsburg by Wahnschaffe, Ref. (3), calcined dolomite was also used and produced SO_2 removal efficiencies of 75% to 90% at a stoichiometric ratio of 1:1. However, as discussed by Kruel and Juntgen, the Wolfsburg test results are suspect due to the possibility of sorbent deposits in the sampling probes. This disturbance occurred during initial testing in their own bench-scale unit, and is indicated in the Wolfsburg tests by the fact that the desulfurization continued or increased when the injection was interrupted or stopped.

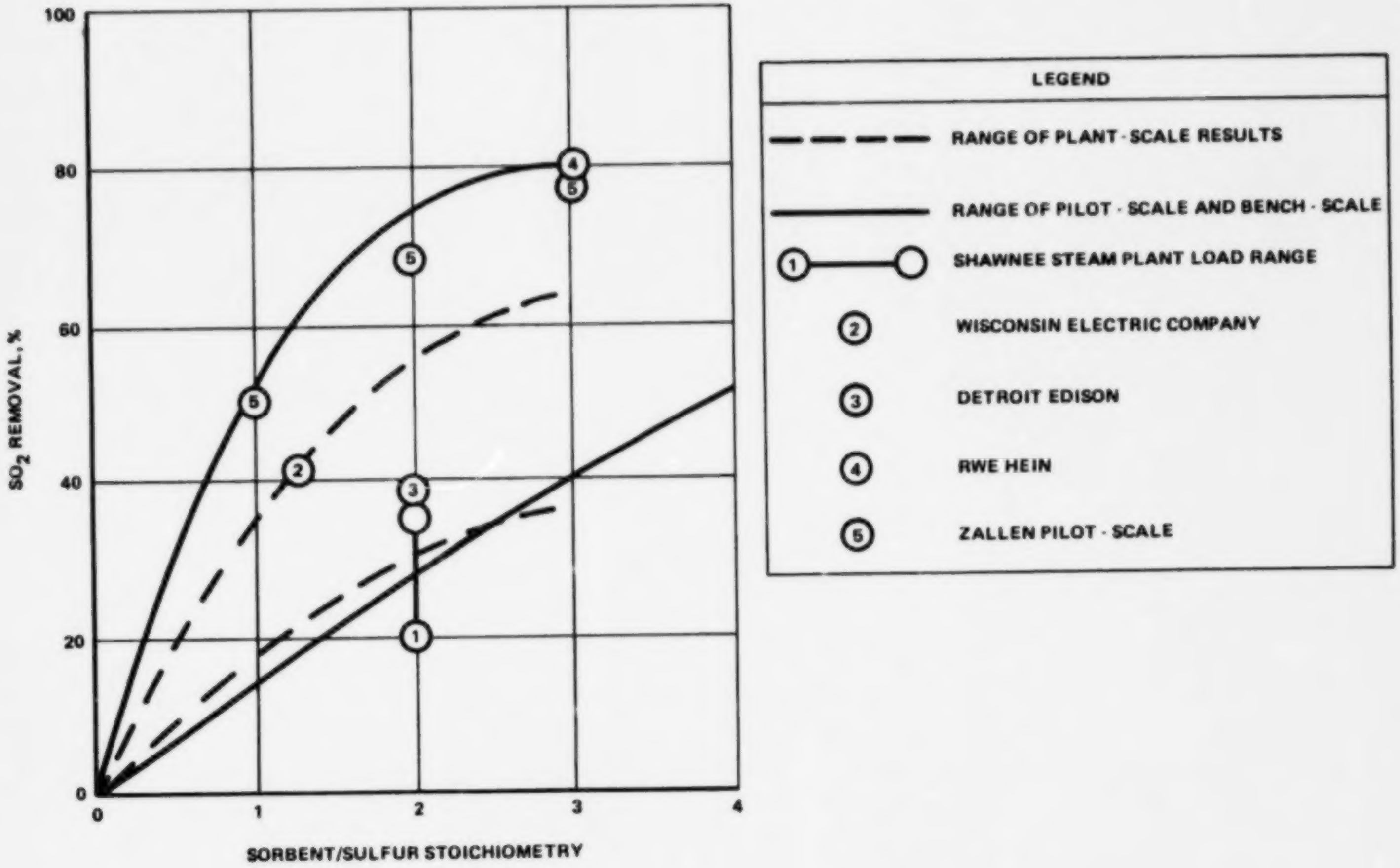


Figure 9. Range of pilot-scale and bench-scale studies (Ref. 17).

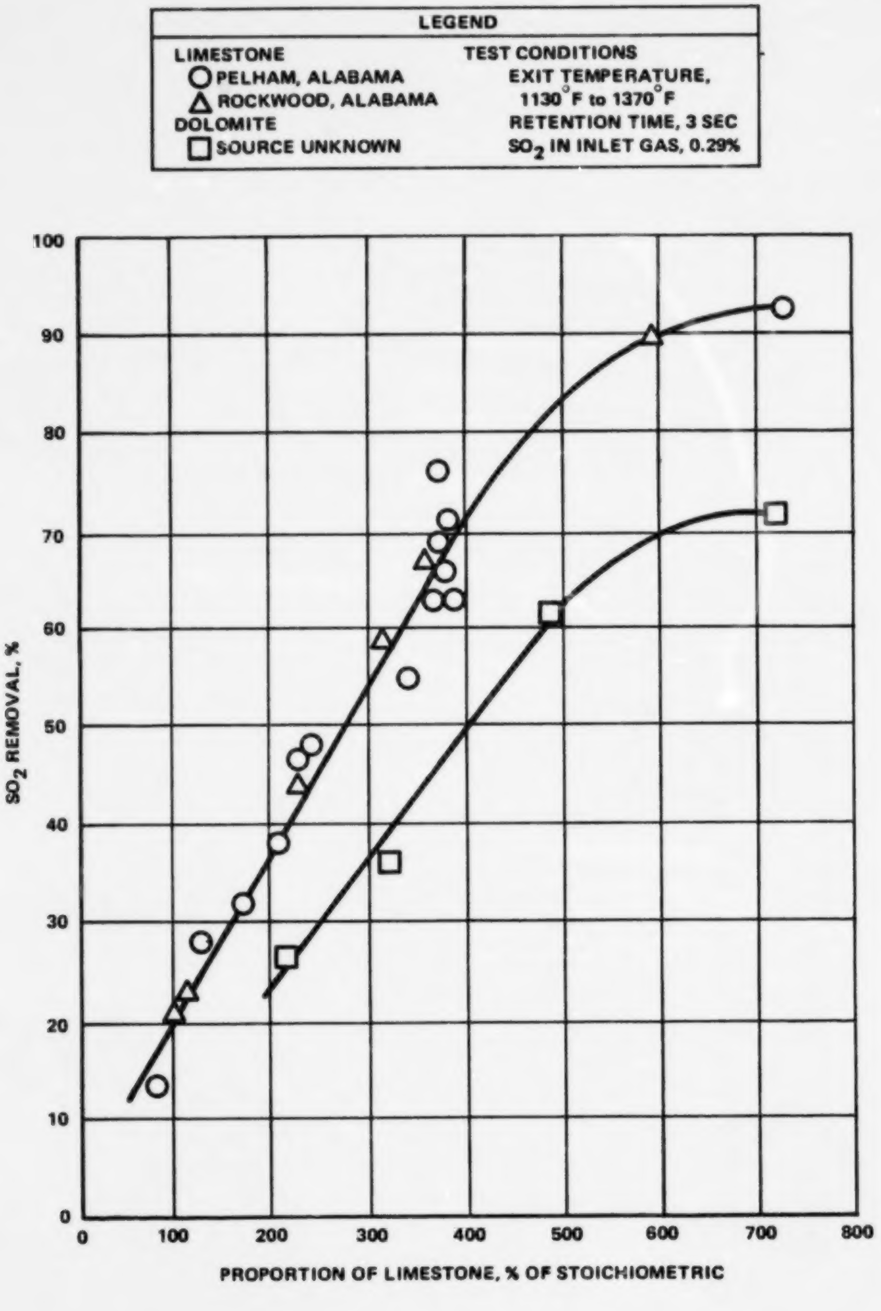


Figure 10. Effect of limestone proportion on removal of SO_2 from air-SO_2 mixture (Ref. 27).

In 1966, tests by Combustion Engineering at Detroit Edison using dolomite and limestone produced SO_2 removal efficiencies of 21% to 28% at a stoichiometric ratio of 1:1. These tests were conducted in one side of a 325 MWe twin furnace burning coal with a 2.5% to 3.5% sulfur content.

For this study (EPA Contract 68-02-3915), a Ca/S ratio of 2:1 was used for high sulfur coals and a Ca/S ratio of 4:1 was used for low-sulfur coals. These ratios were selected after considering several trade-offs between incremental SO_2 removal versus the penalties associated with increased sorbent requirements. For low-sulfur coals, tests have shown that little additional SO_2 removal is achieved at Ca/S levels above 4:1. The added requirements for pulverizing and transporting additional sorbent, plus the increased back-end requirements, more than offset the benefits of added SO_2 removal. For high sulfur coal, the major consideration is solids carryover. At a Ca/S ratio of 2:1, for a 3% sulfur coal, the solids loading would be 150% greater than without sorbent injection (compare with 50% greater loading for a 0.5% sulfur coal at 4:1 Ca/S). This solids loading increase will have an adverse effect on ESP performance. Increased sootblowing will be required to remove the (anticipated) increased deposits in the convective sections.

Zentgraf, referenced by TVA, Ref. (27), carried out plant-scale injection tests at the Kellerman plant of Steinkohlen - Electrizitat AG (STEAG) in Germany. The coal had a sulfur content of about 1.5%. Various materials were injected at different stoichiometries and injection locations (and therefore injection temperatures). These results are shown in Figure 11.

In tests conducted at a BASF power plant in Ludwigshofen, Germany, Goldschmidt reported on tests $Ca(OH)_2$ and dolomite, Ref. (27). Retention time in the boiler was relatively long, 4 to 5 seconds from the 1000°C point to the air heater. Tests were conducted with both coal and oil firing, as shown in Figures 12 and 13. It is interesting to note the higher removal efficiencies with oil firing. This is in line with recent observations by EER and others on the deactivating effect that coal ash appears to have on sorbent activity.

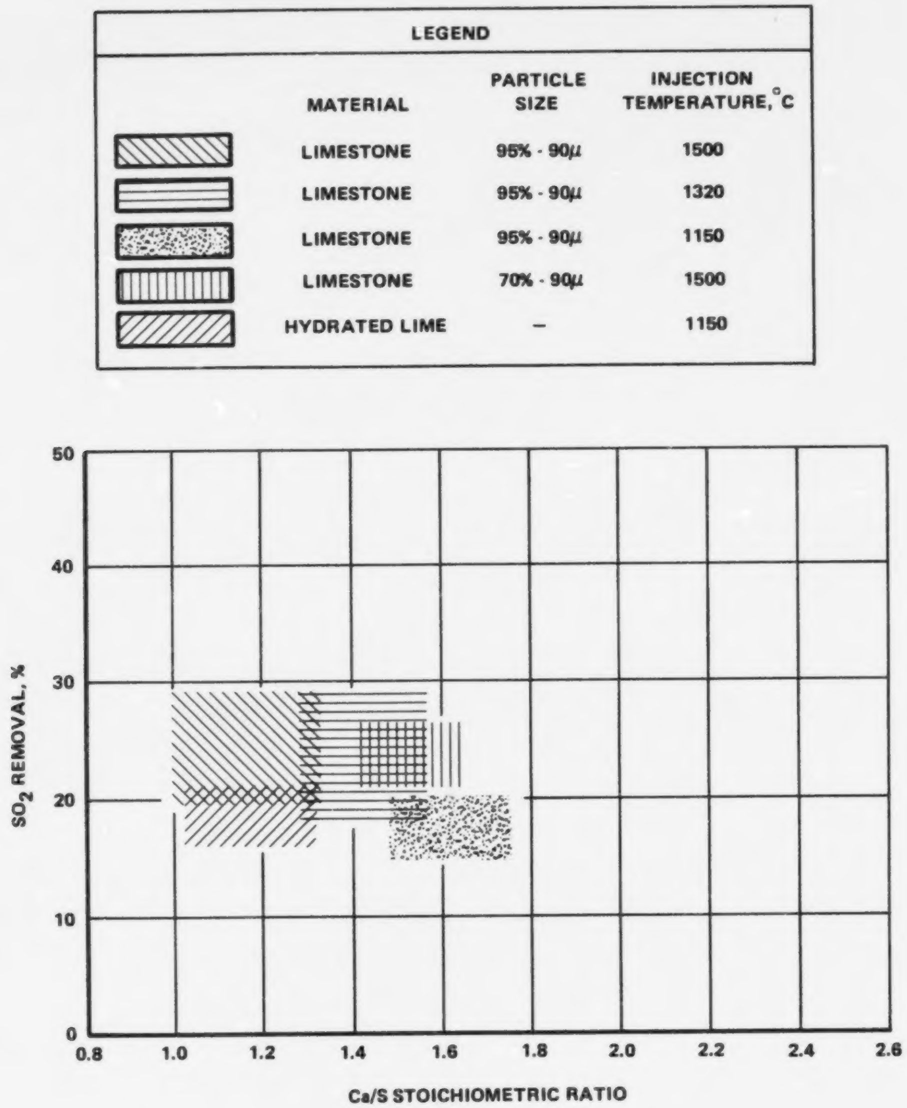


Figure 11. Removal of SO_2 by injection of limestone and lime (Ref. 27).

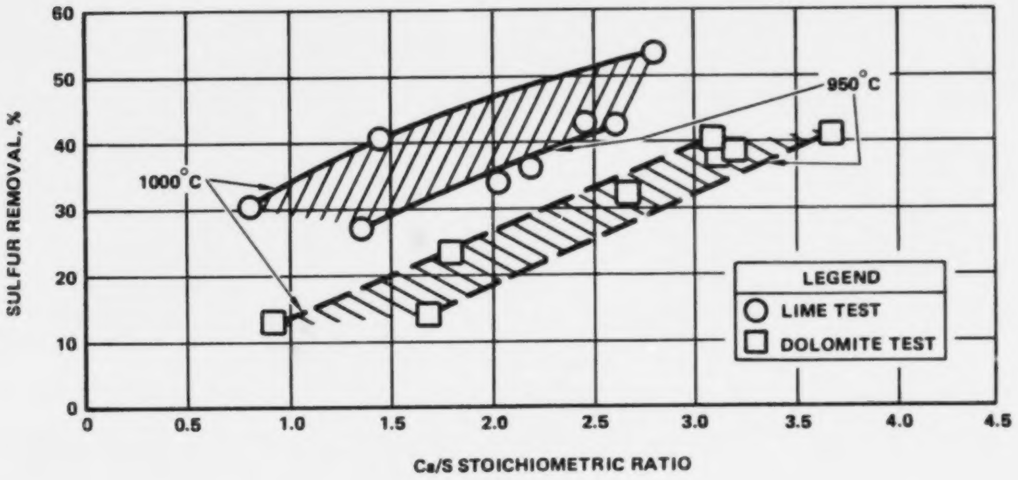


Figure 12. Absorption of SO_2 by hydrated lime and hydrated dolomite injected into a coal-fired boiler (Ref. 27).

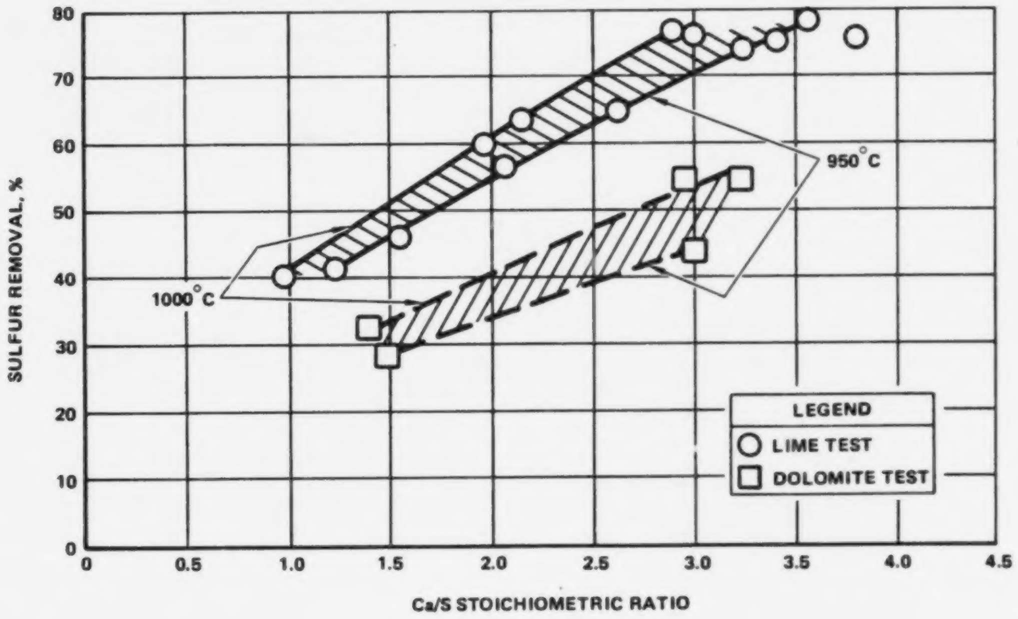


Figure 13. Absorption of SO_2 by hydrated lime and hydrated dolomite injected into an oil-fired boiler (Ref. 27).

In more recent tests, Hein and Glaser, Ref. (15), have reported results from a 4-corner tangentially fired brown coal boiler with a capacity of 175 tons steam/hr (60 MWe). Coal and sorbent are mixed in the milling/drying circuit and fed through the burners. $Ca(OH)_2$, $CaCO_3$, CaO, and dolomite were tested, the results of which are shown in Figure 14. In terms of SO_2 removal efficiency at equivalent stoichiometries for the four sorbents, performance was, from best to worst: $Ca(OH)_2$, dolomite, $CaCO_3$, and CaO. With $CaCO_3$, typical performance results showed removal efficiencies of about 40% for a stoichiometric ratio of 1:1 at an SO_2 concentration of 200 mg SO_2/m^3. Removal efficiency decreased at lower SO_2 concentrations with equivalent sorbent stoichiometry.

The concept of using sorbent injection with low-NO_x burners operating at reduced flame temperatures has been tested on pilot-scale units and small utility steam generators. Steinmuller, Ref. (23), conducted pilot tests with staged mixing burners (SM-burner) using heavy fuel oil with a sulfur content of 1.8%, and bituminous coal with a sulfur content of 2.4%. The sorbent used was slaked lime, $Ca(OH)_2$. At a stoichiometric ratio of 3:1, 87% removal was achieved using coal and 68% with fuel oil. These figures also demonstrate a higher efficiency at higher SO_2 concentrations, although different fuels were used in this case.

Testing is currently in progress by Energy Environmental Research (EER), Ref. (17), using low-NO_x burners and a boiler simulator furnace (BSF) system. Results from the low-NO_x burners (50 M Btu/hr, EPA and Steinmuller burners) show that 60% SO_2 removal is attainable at a stoichiometric ratio of 2:1. Performance has been system dependent, depending on fuel type, burner conditions, and firing rate.

The effect of calcium to sulfur stoichiometry, residence time, reaction temperature and chemical composition of the gas mixture was studied by Mitsubishi Heavy Industries, Inc. (MHI), Ref. (18), in the 1-inch diameter tube reactor shown in Figure 15. Gas mixtures simulating flue gas from a pulverized coal-fired boiler passed downward through an electrically heated

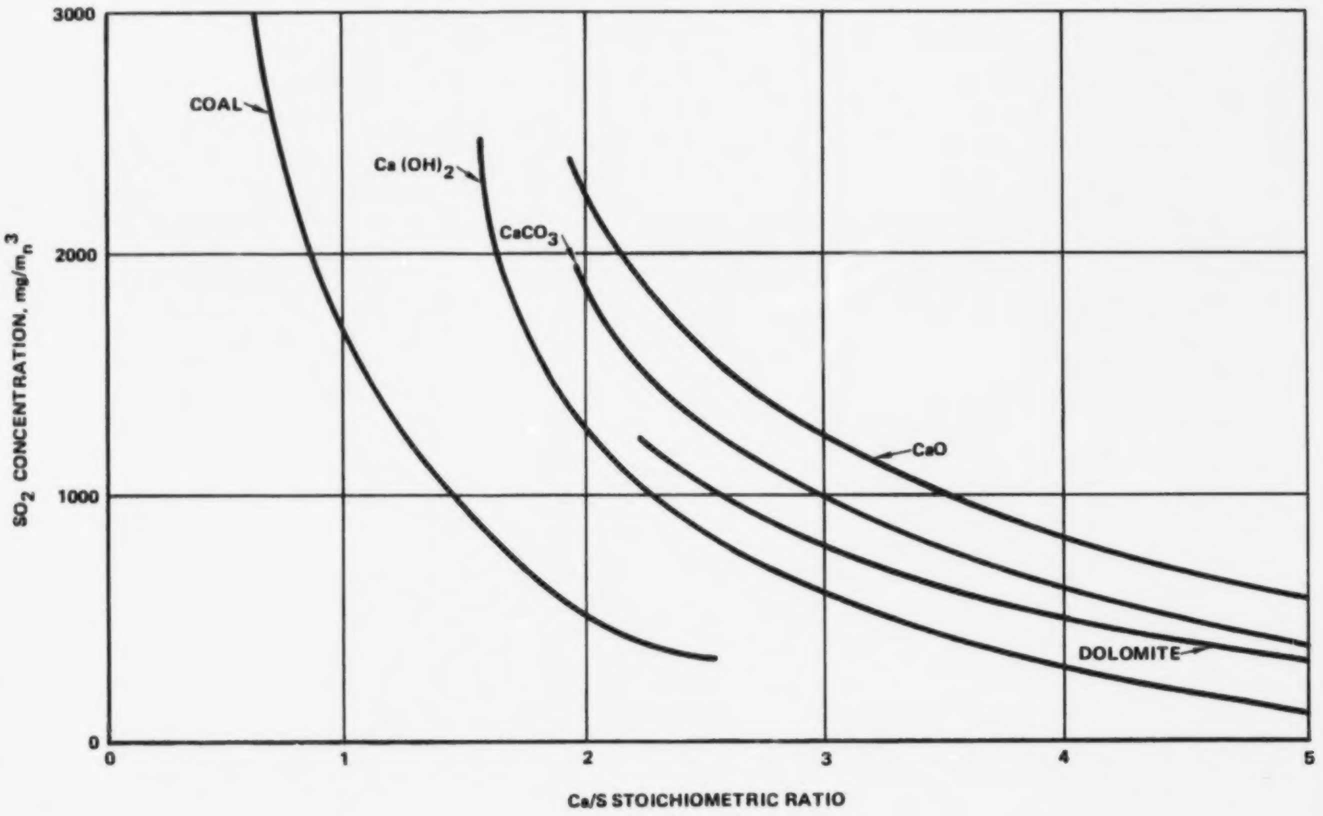


Figure 14. Relative SO_2 removal effectiveness for various sorbents (Ref. 15).

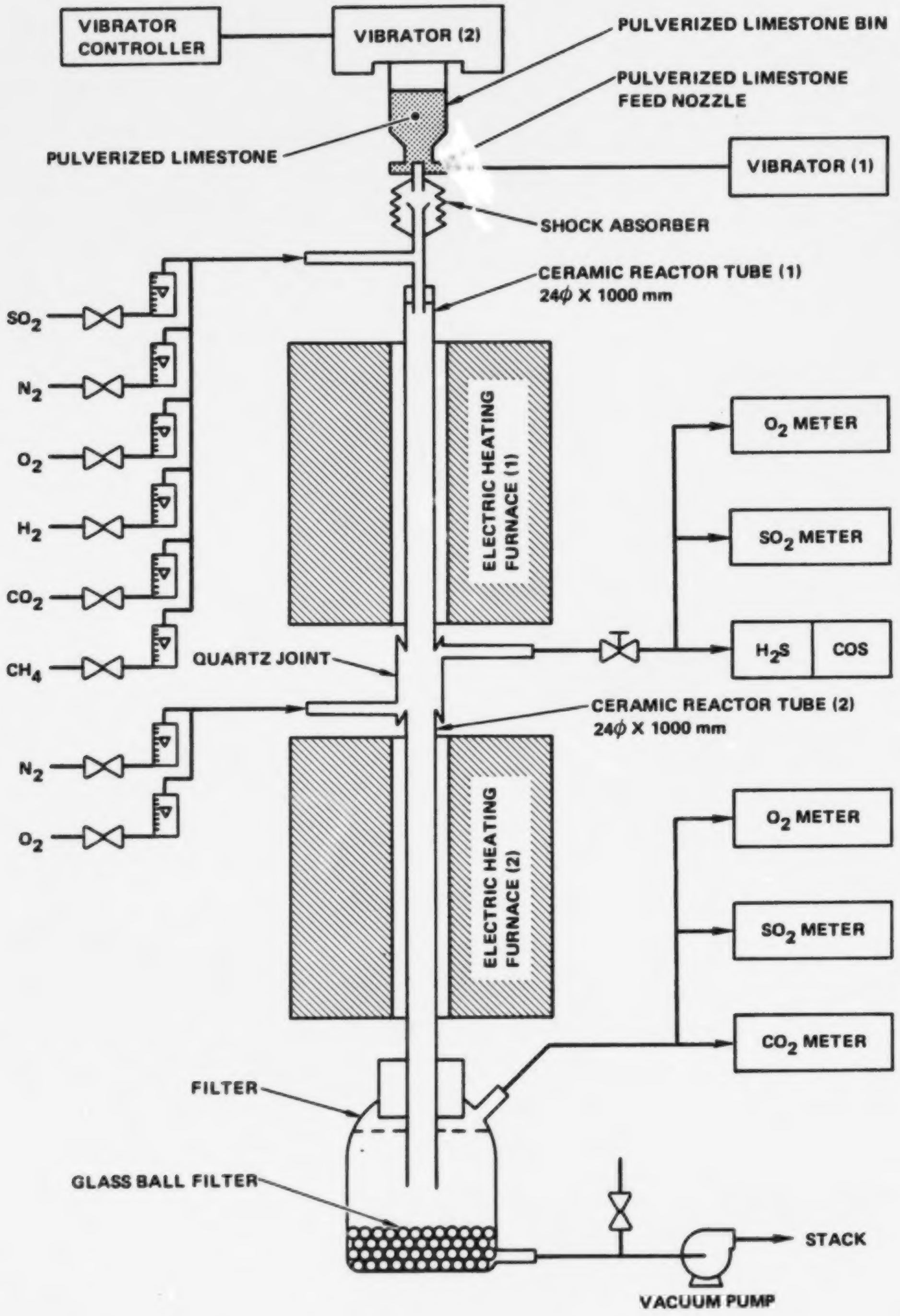

VIBRATOR CONTROLLER
VIBRATOR (2)
PULVERIZED LIMESTONE BIN
PULVERIZED LIMESTONE FEED NOZZLE
PULVERIZED LIMESTONE
VIBRATOR (1)
SHOCK ABSORBER
CERAMIC REACTOR TUBE (1) 24φ X 1000 mm
SO2
N2
O2
H2
CO2
CH4
ELECTRIC HEATING FURNACE (1)
O2 METER
SO2 METER
H2S
COS
QUARTZ JOINT
CERAMIC REACTOR TUBE (2) 24φ X 1000 mm
N2
O2
ELECTRIC HEATING FURNACE (2)
O2 METER
SO2 METER
CO2 METER
FILTER
GLASS BALL FILTER
STACK
VACUUM PUMP

Figure 15. Reactor tube test facility for SO_2 removal by pulverized limestone (Ref. 18).

ceramic tube. Pulverized limestone was fed from above the tube at rates controlled by adjustments to the vibrator. As shown in Figure 16, SO_2 capture increased with increasing stoichiometry and residence time. At a stoichiometric ratio of 3:1 and a residence time of 1.54 sec., greater than 90% SO_2 removal was measured.

4.2.5 Distribution and Point of Injection

Many different schemes have been tried for injecting sorbent into boilers for sulfur removal. From these tests, several criteria for injecting sorbent to achieve the most beneficial results have been determined: (1) maximize sorbent particle residence time in the "critical" temperature region (between 2250°F and 1650°F for limestone), where the calcium/sulfur reaction to form calcium sulfate is most favorable; (2) minimize the time the sorbent particle is subject to hardburning (above 2700° for limestone); (3) maximize the particle surface area; and, (4) maximize mixing between the sorbent particles and the flue gas.

Because of the short retention time in most utility boilers, it is important for the sorbent to be distributed rapidly over the boiler cross section in order to contact all areas of the gas stream. If time is lost in achieving the proper distribution, then some portions of the stream will not have enough time for good absorption after the sorbent becomes available.

Distribution is a difficult problem because of the very large cross section of modern boilers, the rapid flow of the gas, and the small size of the particles. In most injection systems, the particles are carried in a body of injection gas, so that the problem becomes one of fluid dynamics--the mixing of one body of gas with another.

The number of injection ports is an important consideration for sorbent injection. It is apparent from recent testing that multiple injection locations will be required to obtain the proper amount of mixing within the available residence time. The gas velocity through the boiler approaching the

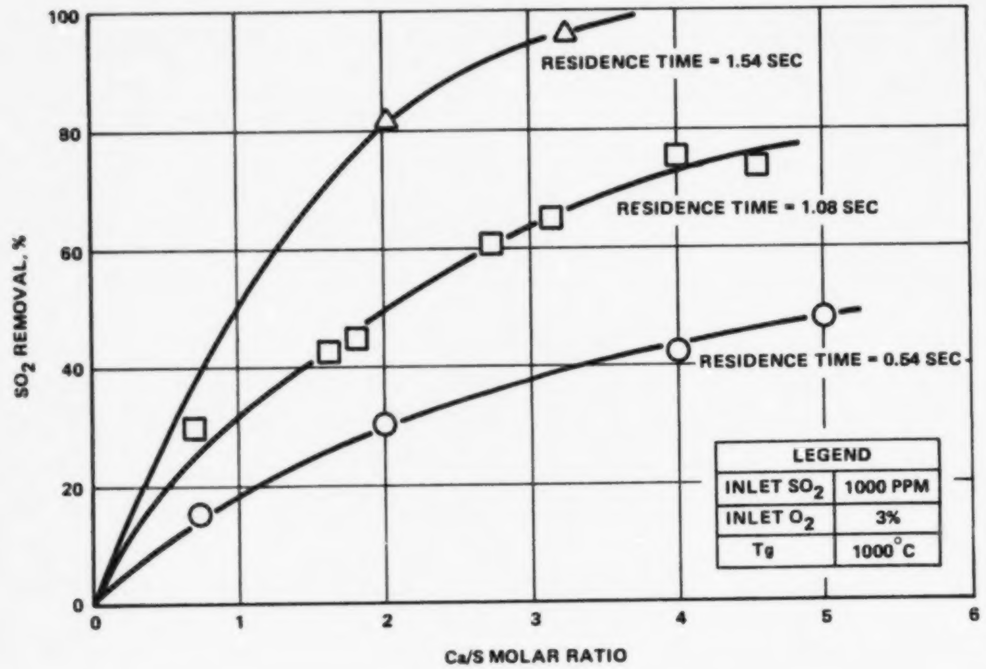


Figure 16. Relation between SO_2 removal and Ca/S molar ratio (Ref. 18).

suggested injection point of superheat panels varies from 35 to 50 fps. The sorbent will have to be injected at such a velocity and angle as to maximize the contacting with the flue gas during the roughly 1.5 sec residence time in the critical temperature range.

In the testing at TVA-Shawnee, more effort was expended compared to other investigations with regard to characterizing and attempting to improve sorbent distribution. Their conclusion was that furnace gas turbulence is the variable which has the greatest effect on sorbent distribution. Attempting to impose outside control over sorbent distribution had little effect. In other words, variables such as limestone injection velocity or angle did not significantly affect sorbent distribution over the range tested. Injection velocities ranged from 60 to 130 ft/sec.

Recent EER tests, Ref. (17), were conducted at two different velocities while injecting limestone through the tertiary air ports of the LWS furnace. Tests were run with both Utah coal (0.87% S) and Indiana I coal (2.57% S) at velocities of 80 ft/sec and 100 ft/sec. As shown in Figure 17, a significant improvement occurred at the higher injection velocity. However, the question still remains as to whether this effect would be seen in full-scale boilers.

In addition to mixing considerations, the location of sorbent injection also determines the temperature encountered, gas mixture (reducing or oxidizing), and particle residence time. These factors can influence both calcination of the sorbent and the reactions between sulfur species in the gas phase and the sorbent. Effects of residence time and temperature have largely been covered in the preceding section with respect to sorbent calcination. The effects of these variables on the absorption reactions, and ultimately on the overall quantity of absorption and sorbent utilization, will be discussed in the following section.

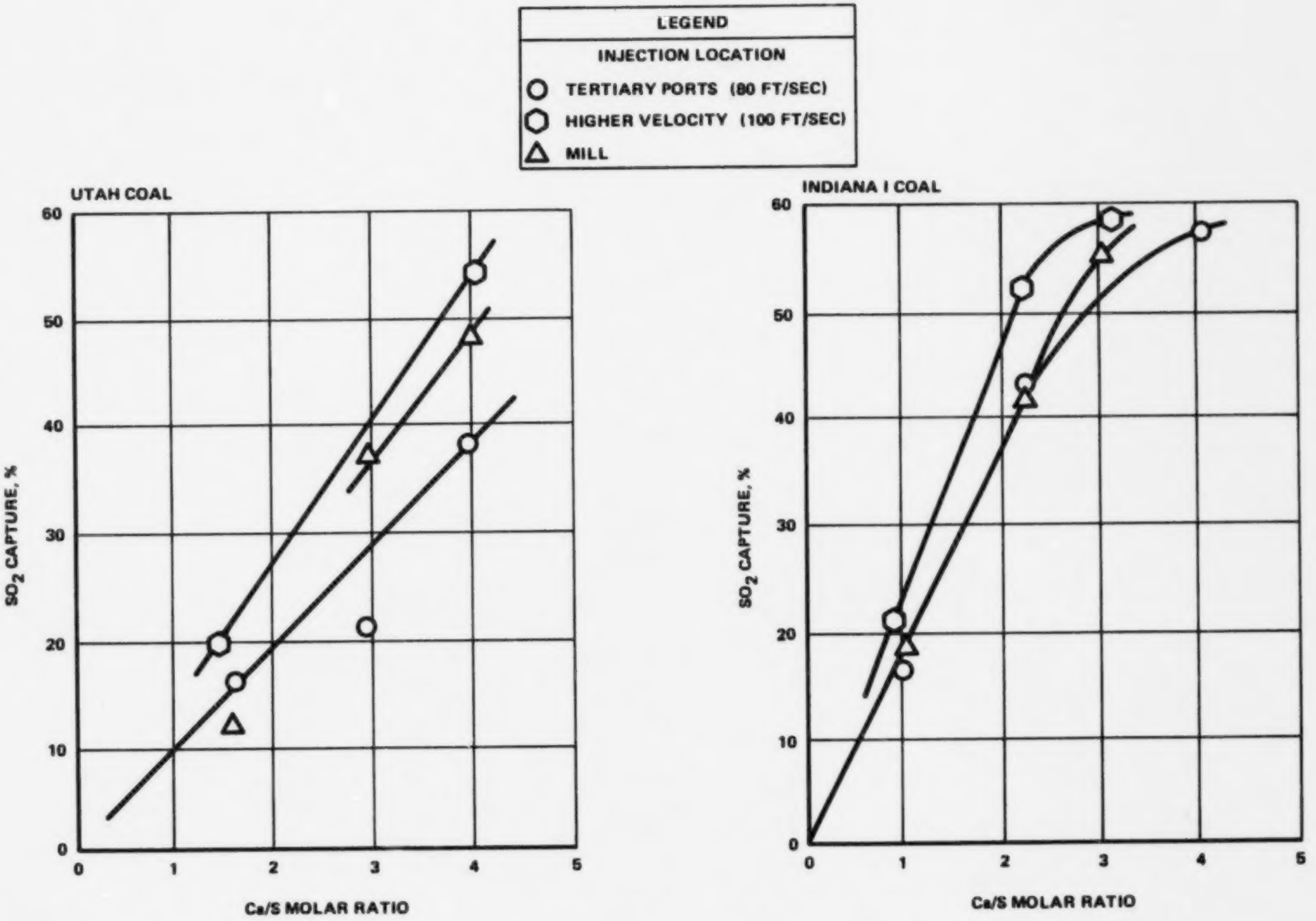


Figure 17. EPA DMB testing: SO_2 capture vs. Ca/S molar ratio for Utah coal and Indiana coal on the large water-tube simulator furnace ($CaCO_3$ sorbent, 70 x 10^6 Btu/hr) (Ref. 17).

4.3 Sorbent/Furnace Environment Interaction

Optimum absorption of SO_2 by sorbent injection requires a balance between the rate of reaction of the sorbent with SO_2 and the rate of decomposition of the products formed. Since rate of reaction increases with temperature, higher gas temperature is beneficial. However, the operating temperature must be below the decomposition temperature of the products involved. Other factors which influence the capture of SO_2 are the composition of the flue gas and the composition of the coal ash. This section includes a discussion of the proposed gas/sorbent chemical reactions, and the effects of time/temperature, flue gas composition, and coal ash composition upon SO_2 capture.

4.3.1 Proposed Reaction Steps

In the combustion process under normal oxidizing conditions, sulfur in the fuel is converted primarily to SO_2, while a small portion is oxidized to SO_3. Under reducing conditions, H_2S and COS are formed, and these compounds may exist at least initially in the normal combustion process as well. According to Wickert, referenced by Chughtai, Ref. (23), in the case of SO_2 absorption by calcium alkali at flue gas temperatures up to 400°C, the process occurs in two steps, as follows:

$$CaO + SO_2 \rightarrow CaSO_3$$

Calcium sulfite ($CaSO_3$) is eventually oxidized according to:

$$CaSO_3 + 1/2\ O_2 \rightarrow CaSO_4$$

At higher flue gas temperatures, $CaSO_3$ is not stable and the absorption process occurs by:

$$4\ CaO + 4\ SO_2 \rightarrow 3\ CaSO_4 + CaS$$

and,

$$CaS + 2\ O_2 \rightarrow CaSO_4$$

SO_3 will react directly with CaO by:

$$CaO + SO_3 \rightarrow CaSO_4$$

In the case of H_2S or COS in the flue gas, the steps would be:

$$H_2S + CaO \rightarrow CaS + H_2O$$

or,

$$COS + CaO \rightarrow CaS + CO_2$$

and,

$$CaS + 2\ O_2 \rightarrow CaSO_4$$

4.3.2 Time/Temperature Effects upon SO_2 Capture and Product Decomposition

The capture of SO_2 by the sorbent is dependent upon gas/solid chemical kinetics and the residence time at temperatures favorable for SO_2 capture. Product decomposition, on the other hand, can be predicted by thermodynamic calculations. In this section the temperature regimes of product decomposition are discussed first, followed by a discussion of SO_2 capture kinetics studies.

In the thermochemical study by Battelle it was determined that, at atmospheric pressure, SO_2 removal by CaO is limited to temperatures below 2250°F, and SO_2 removal by MgO is limited to temperatures below 1550°F. Similar information is given by Flament, Ref. (24). If the partial pressure of SO_2 is 500 ppm, $MgSO_4$ begins to decompose at 1600°F. Similarly, $CaSO_4$

begins to decompose at 2200°F, and $CaSO_3$ begins to decompose at 1332°F. The low decomposition temperature of $CaSO_3$ rules out the existence of $CaSO_3$ at average boiler temperatures.

Battelle evaluated the rate of reaction of $CaSO_3$ with SO_2 in the Dispersed Phase Reactor (DPR) at temperatures between 1500°F and 2150°F. Results of these tests are shown in Figure 18 in an Arrhenius plot indicating an activation energy of 28 kcal/mole. It should be noted that both calcination and reaction of CaO with SO_2 were occurring simultaneously in the reactor. According to Battelle, a similar activation energy (25 kcal/mole) was found by Ishihara in the temperature range 1800°F to 2200°F. It was concluded that the magnitude of the apparent activation energy suggests that the reaction is not diffusion controlled.

According to work by Wickert, referenced in Kruel and Juntgen, Ref. (4), limestone and dolomite have their highest reaction rate in the temperature range 1472°F to 1832°F. For CaO, the rate fell off sharply above 2192°F, while for dolomite the rate decreased rapidly above 1832°F. In other testing and observations, Zentgraf reported 2100°F as the optimum for injecting calcium hydroxide.

In the TVA full-scale testing, it was observed that a significant increase in SO_2 removal occurred when the sorbent was injected at the higher furnace elevation where the gas temperatures were lower. Temperatures at the various injection levels were estimated to be in the 2400°F to 2900°F range. However, in general, SO_2 removal was low--approximately 20% at a stoichiometry of 2:1. In brown coal tests in Germany, Hein and Glaser report excellent SO_2 removal results (approximately 75% at Ca/S = 2:1) in a furnace where temperatures are normally below 2100°F.

Recent work by EER also demonstrated the beneficial effect of lower gas temperatures. Figure 19 is a plot of SO_2 removal vs. Ca/S molar ratio and temperature vs. heat extraction and residence time in the furnace. The data show that additional cooling improved SO_2 removal by increasing residence time in the critical 2200°F to 1400°F temperature zone.

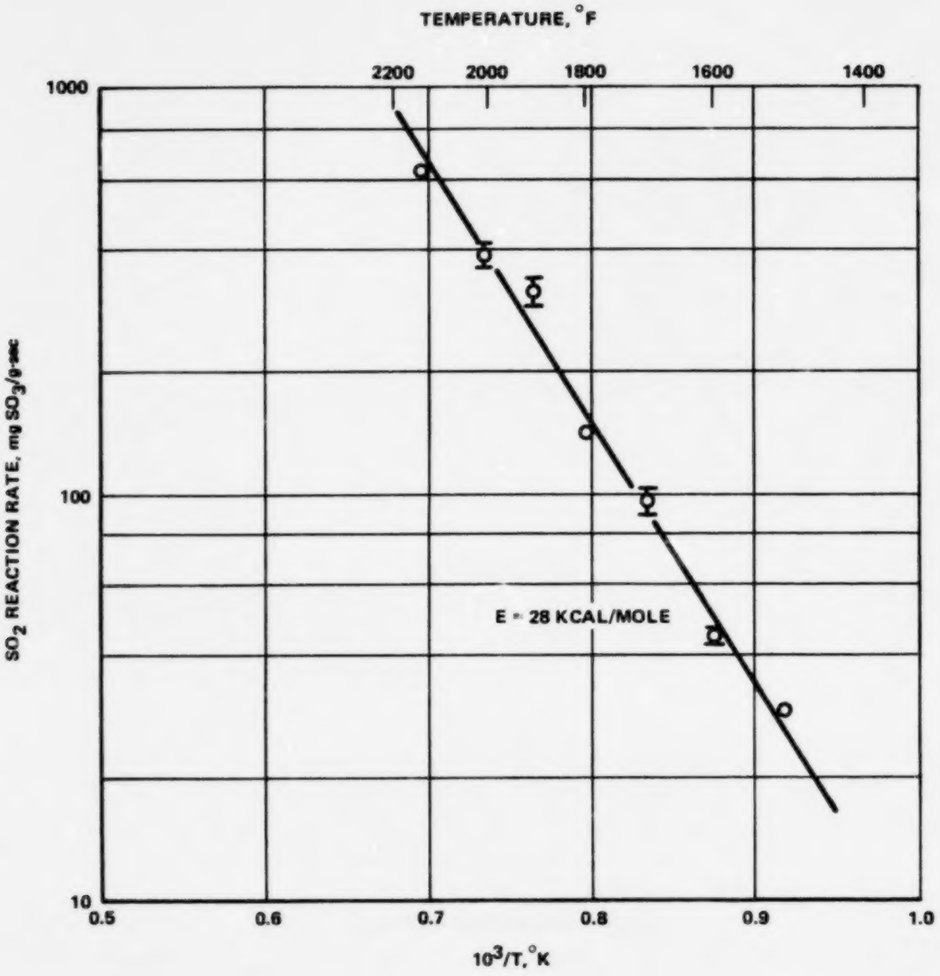
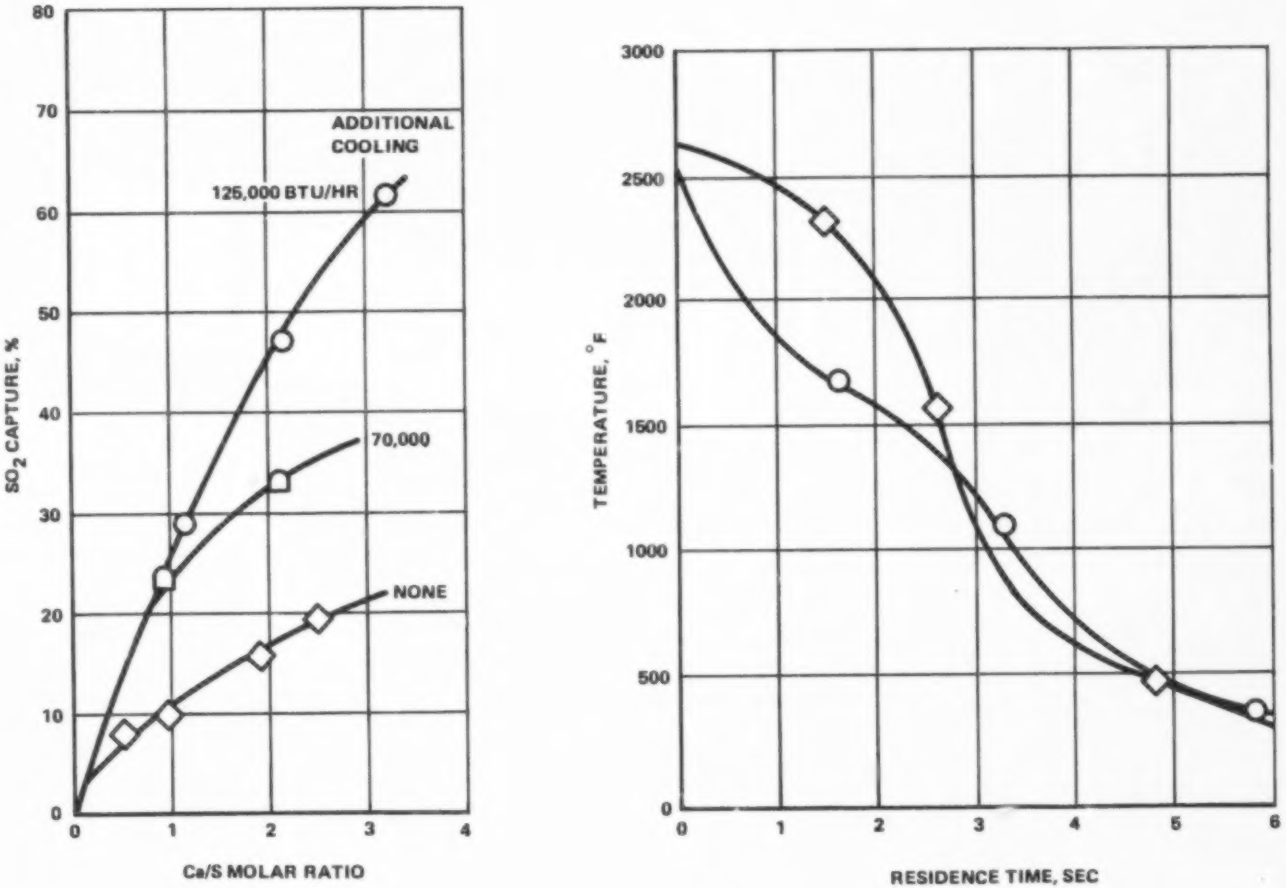


Figure 18. Temperature dependence of initial SO_2 reaction rate in reactor for BCR-1359 (Ref. 5).



SORBENT - $CaCO_3$, 45-3
LOCATION - TERTIARY AIR
ALL DATA PORT NO. 3

DATES	6/25	COOLING AS SHOWN	
	7/1	SR_P	0.2
	7/29	SR_B	0.6
RATE	800K	SR_T	1.2
COAL	IND.	OTHER	–
% S	–		

Figure 19. EPA DMB testing: effect of increased cooing on SO_2 capture and gas residence time ($CaCO_3$ sorbent, Indiana coal) (Ref. 17).

Data obtained in the MHI tube reactor, Figure 20, indicate that the optimum temperature for SO_2 capture is between 950°C and 1100°C (1700°F to 2100°F). As expected, SO_2 capture increases with increasing residence time.

4.3.3 Effect of SO_2 Concentration

Initial DPR testing and micro balance experiments at Battelle suggested a first order dependence between reaction rate and SO_2 concentration. However, later tests indicated that the rate of reaction of limestone with SO_2 was essentially independent of the SO_2 concentration. Another variable, loading or lime utilization, was determined to affect the apparent order of reaction. These additional tests, shown in Figure 21, illustrate that the overall reactivity is proportional to the square root of the SO_2 concentration. Battelle states that this result is in agreement with the B&W work if the B&W data is converted to the same basis. The end result is that both Battelle and B&W data indicate a reaction order of about 1.5.

In tests at EER, higher overall removal rates were also observed at higher SO_2 concentrations in the flue gas. Two levels of SO_2 concentration were examined, approximately 1100 ppm and 2500 ppm. Results from these tests are shown in Figure 22. Tests at MHI, shown in Figure 23, further confirm that SO_2 removal increases at increased SO_2 concentration.

4.3.4 Effects of O_2 Concentration, Gas Composition, and Oxidation Catalysts

The overall reaction for absorption of SO_2 by calcined limestone is:

$$CaO + SO_2 + 1/2\ O_2 \rightarrow CaSO_4$$

Oxygen for this reaction is supplied by the excess air in the flue gas. In the absence of sorbent injection, oxidation of SO_2 takes place only to a limited extent, about 0.5% to 1.0% of the SO_2 is oxidized in the boiler. With sorbent injection, oxidation of the sulfur absorbed is essentially complete, resulting in a product material of $CaSO_4$. However, the mechanisms and rates of oxidation are not completely known. The equilibrium conversion of SO_2 to SO_3 is not higher at the furnace temperatures involved, but reaction of

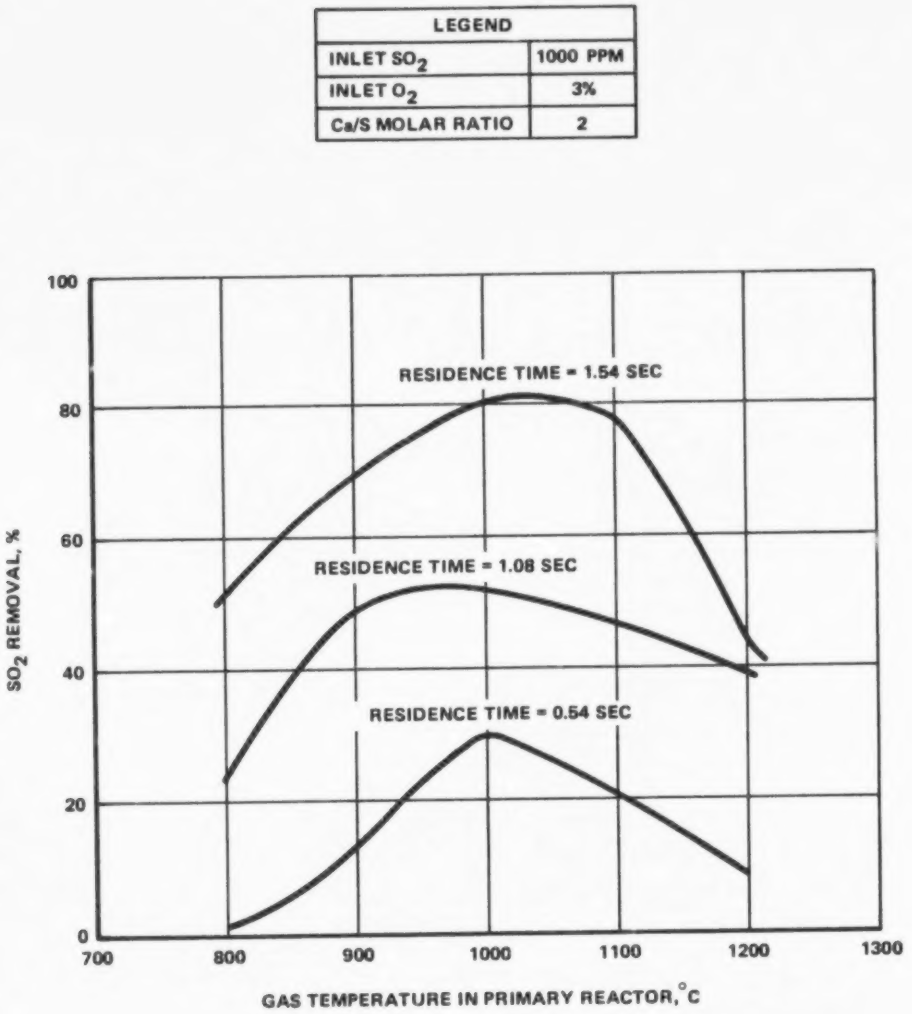


Figure 20. Relation between SO_2 removal and gas temperature (Ref. 18).

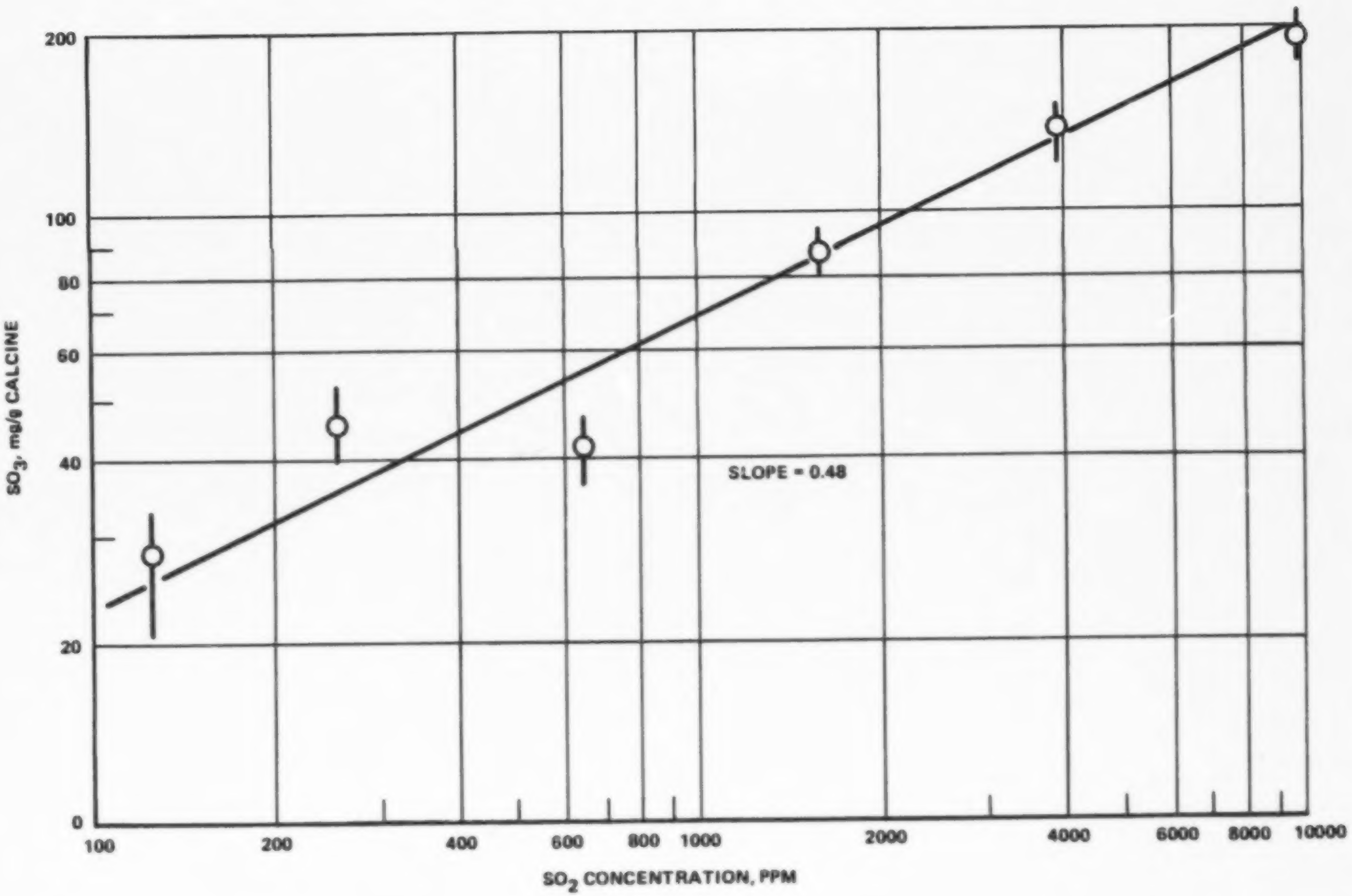


Figure 21. Effect of SO_2 concentration on overall sulfur pickup (Ref. 26).

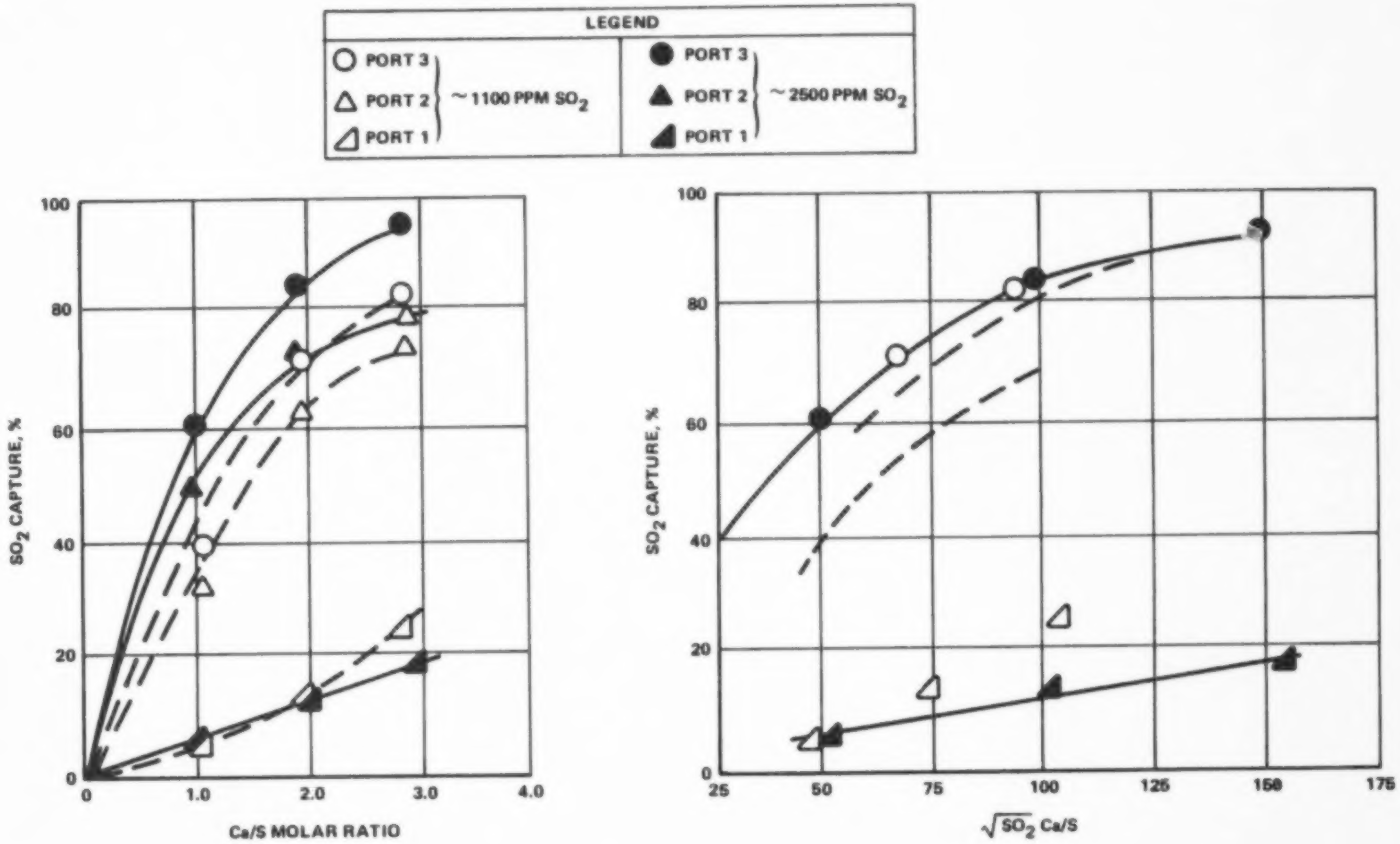


Figure 22. EPA DMB testing: increased SO_2 removal rate with higher SO_2 concentration (Ref. 17).

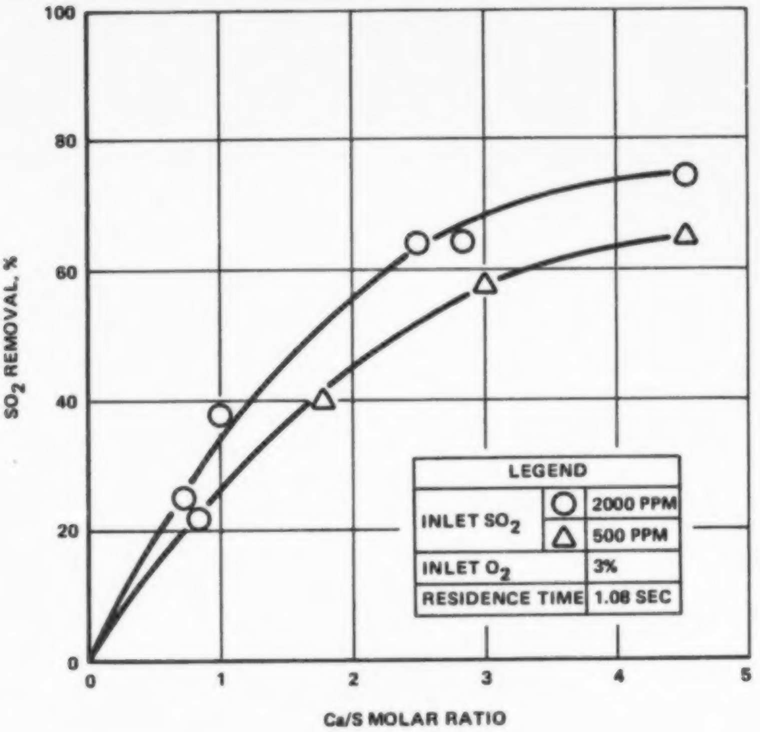


Figure 23. Relation between SO_2 removal and Ca/S molar ratio with different SO_2 concentration (Ref. 18).

SO_3 with CaO is presumably a fast reaction and should promote oxidation of SO_2 by removing the reaction product. It is also a possibility that $CaSO_3$ may form first and then oxidize, but the low decomposition temperature of $CaSO_3$ would require a rapid oxidation step. In an effort to provide some answers to the above questions, investigators have examined the effects of flue gas oxygen content, nitrogen oxide content, and potential oxidation catalysts present in or added to sorbents.

In tests conducted on Battelle's micro balance, selected limestone samples were calcined and reacted in the absence of air. Based on these tests, it was hypothesized that $CaSO_3$ is formed first, especially towards the inner regions of the particle.

In the pilot-scale tests conducted by B & W, oxygen content in the flue gas was varied from 3% to 0.2%. SO_2 removal for these tests decreased from 14% to 2% as oxygen concentration decreased.

At TVA, O_2 concentration was varied in a higher range (1.0% to 10.0%) than the B & W tests, and no effects on SO_2 removal were observed.

In order to evaluate the effect of O_2 concentration and reducing gas mixtures upon SO_2 capture, MHI compared sulfur capture from gases containing SO_2 and 3% O_2 to sulfur capture from various other gas mixtures. Figure 24 shows that the reactivity of H_2S with limestone is about one-half of that of SO_2 with limestone in an oxidizing atmosphere. Figure 25 shows that the reactivity of SO_2 with limestone without O_2 is less than the reactivity of SO_2 with limestone and 3% O_2. This is true whether CO is or is not present. The presence of hydrogen (H_2) has an anomalous effect since H_2 removes SO_2 by converting SO_2 into elemental sulfur. This occurred to an almost equal degree whether limestone was or was not fed. Figure 26 shows that the presence of methane in the gas mixture had a slight negative effect on sulfur capture.

With respect to oxidation promoters or catalysts, studies have been made on effects of intrinsic impurities and on the use of additives. One of the more extensive evaluations of possible catalysts was conducted by Battelle in

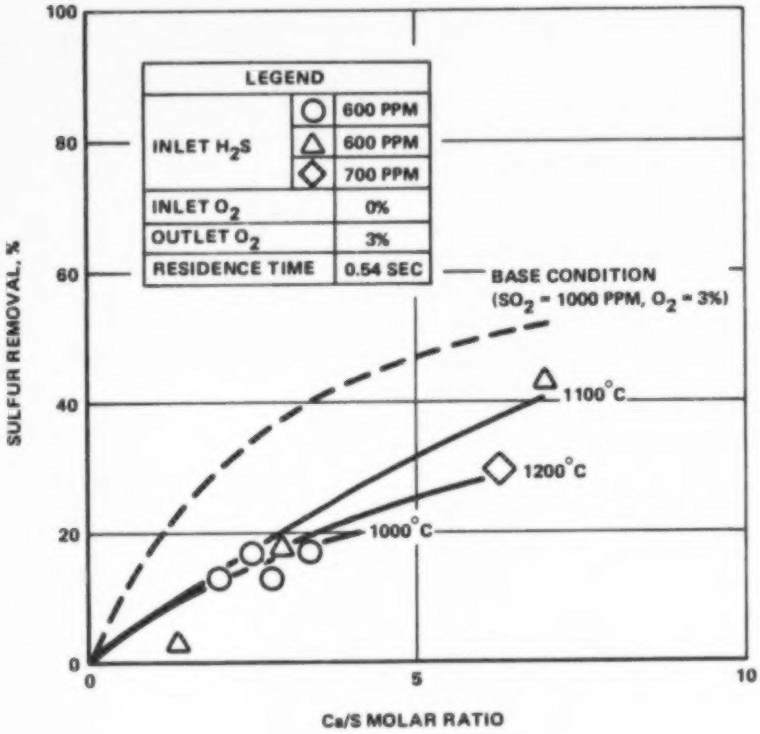


Figure 24. Relation between SO_2 removal and Ca/S molar ratio with 0% O_2 atmosphere (Ref. 18).

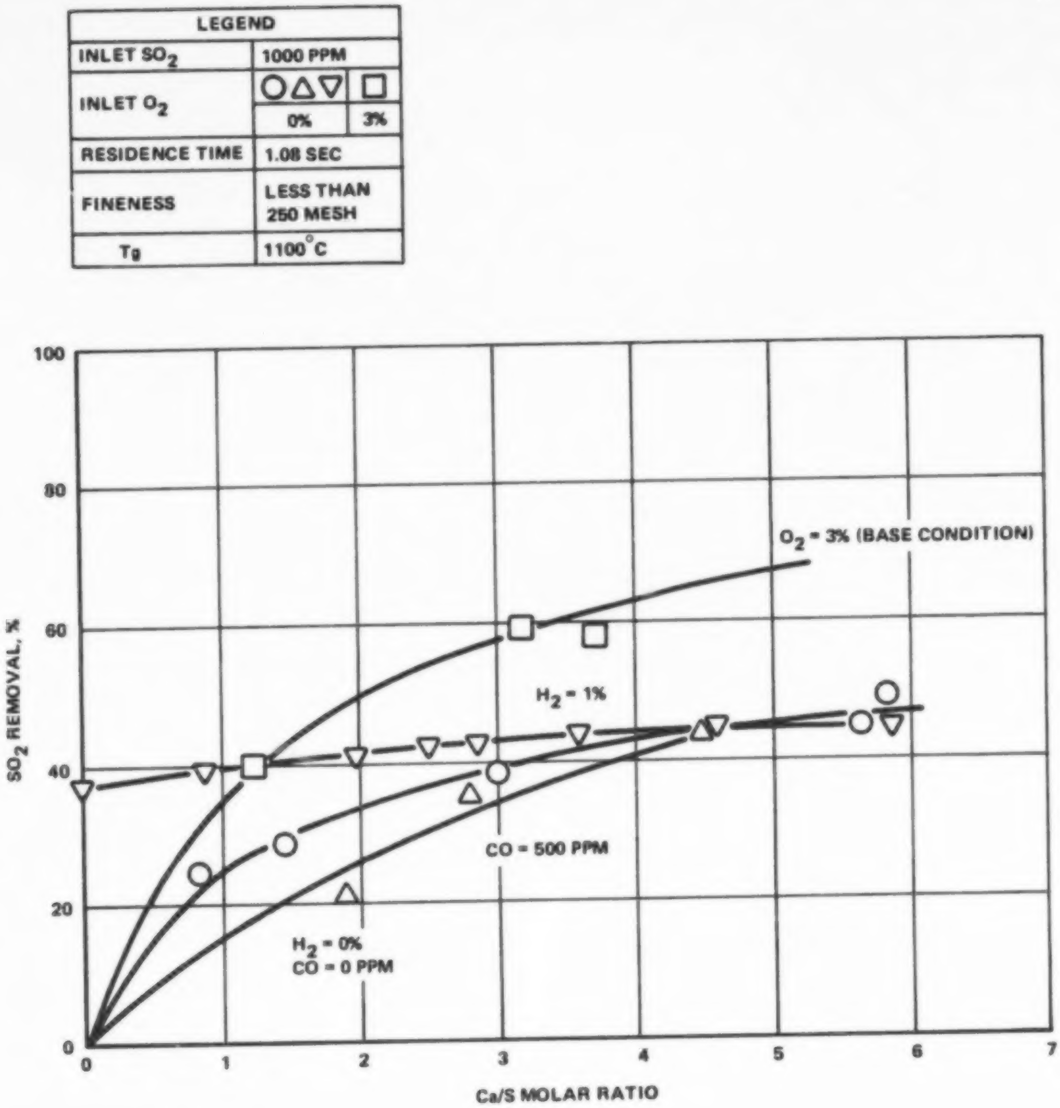


Figure 25. Effect of inlet gas composition on SO_2 removal (Ref. 18).

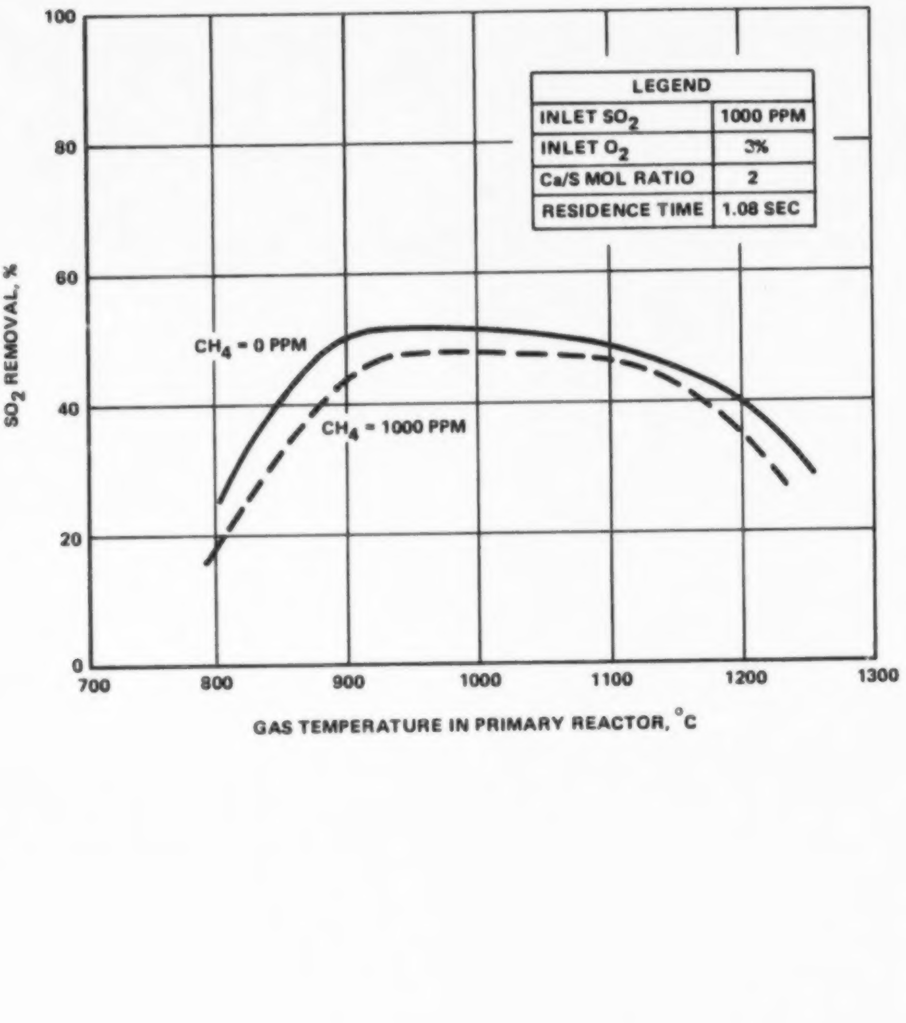


Figure 26. Relation between SO_2 removal and gas temperature (Ref. 18).

the micro balance reactor. Various compounds were added to the limestone before calcining, with the method of addition varying also. A summary of the tests run and the addition method is shown in Table 6. Overall, it was concluded that no change in calcination characteristics occurred, and rates of reaction with SO_2 were not appreciably different compared to untreated stone. It was noticed that glass-forming materials, such as B_2O_3 and P_2O_5, caused a slight reduction in SO_2 absorption rate. No tests or conclusions were made regarding the effects of catalysts on the $MgO-SO_2$ reaction.

In the pilot-scale tests by Ishihara, limestones intrinsically containing over 1% ferric oxide (Fe_2O_3) were about 50% more effective than those containing less than 0.5%.

With respect to MgO, Battelle concluded that reagent grade $MgCO_3$ did not react with SO_2 in their microbalance tests at temperatures of 740°F, 1256°F, and 1530°F. It is significant that these were also relatively long tests, 10 to 15 minutes, compared to seconds in the limestone injection process. In studies by Kruel and Juntgen, conversion of MgO to $MgSO_4$ was dependent on the addition of Fe_2O_3. Amounts in excess of 2.0% did not improve results. Reaction of CaO was not affected by Fe_2O_3 in their experiments.

4.3.5 Ash and Flue Gas Effects

In addition to intrinsic impurities found in limestone and dolomite which may have a beneficial or detrimental effect on SO_2 oxidation, calcination, and ultimate porosity, compounds in coal flyash or flue gas may affect the calcination and absorption process. In early tests documented by Goldschmidt (referenced by TVA), plant scale tests were conducted at a BASF power plant in Germany using coal or oil firing. With $Ca(OH)_2$, SO_2 absorption was considerably better with oil than with coal firing. The better results with oil were attributed to surface deposits of CaO in the boiler, which were affected by slag formation when burning coal.

TABLE 6. PROMOTER PERFORMANCE IN THE MICROBALANCE (REF. 5)

Promoter				Capacity[a], mg Gain per 50-mg Charge
		Concentration on Stone		
Form, as Calcined	Reagent Applied[b]	Equiv per 100 Equiv	Weight Percent	
Fe_2O_3	$Fe(NO_3)_3 \cdot 9H_2O$	5.0	12	10.7
Fe_2O_3	Fe_2O_3, slurry in H_2O	3.9	2	7.2
Fe_2O_3	Fe_2O_3, dry mix	3.9	2	8.6
V_2O_5	Calcium acid vanadate	8.4	3.0	9.2
CuO	$Cu(OAc)_2 \cdot H_2O$	0.5	1.1	8.6
CoO	$Co(OAc)_2 \cdot 4H_2O$	0.5	1.4	10.8
NiO	$Ni(NO_3)_2 \cdot 6H_2O$	0.5	1.6	10.8
Sb_2O_3	Sb_2O_5, slurry in H_2O	2.75	1.7	8.9
MnO_x	$Mn(NO_3)_2 \cdot 6H_2O$	0.5	0.9	11.7
Al_2O_3	$Al(NO_3)_3 \cdot 9H_2O$	0.5	1.3	12.0
Ag	$AgNO_3$	0.5	1.7	11.7
B_2O_3	H_3BO_3	--	5(H_3BO_3)	9.8
P_2O_5	$NH_4H_2PO_4$	--	5($NH_4H_2PO_4$)	9.2
Na_2SO_4	Na_2SO_4	0.5	0.8	7.6
K_2SO_4	K_2SO_4	0.5	1.0	7.6
Na_2CO_3	$NaHCO_3$	0.5	0.9	11.9
Fe_2O_3	$Fe(NO_3)_3 \cdot 9H_2O$	5.0	12	(6.3)[c]
Fe_2O_3	Fe_2O_3, dry mix	3.9	2.0	(6.4)[c]

[a] Reference capacity for untreated BCR-1359 is 12.3 ± 0.5-mg gain per 50-mg stone. No replicates were run. Values probably are correct within several tenths of a milligram.

[b] Aqueous solutions, except as indicated.

[c] Simultaneous calcination and sulfurization at 1380°F to 1420°F.

More recent testing at EER using gaseous versus coal fuels has suggested a deactivation mechanism due to coal combustion products reacting with the calcined limestone. At similar thermal conditions, SO_2 absorption as a function of Ca/S ratio was determined to be much better while burning propane than while firing Indiana #3 coal. These comparisons were made for vicron (limestone), $Ca(OH)_2$, and dolomite. The results are shown in Figure 27. In order to isolate which constituents of coal ash most affect absorption, various materials and compounds were added to vicron. As shown in Figure 28, Al_2O_3 appears to be one of the common and significant compounds causing decreased absorption of SO_2 when injecting vicron.

4.4 Sorbent Preparation and Delivery

Sorbent preparation requirements vary substantially among the various sorbent types. For example, preparation of $Ca(OH)_2$ is more complex and costly than preparation of limestone. Since limestone is the sorbent selected for this study, this discussion will be oriented toward limestone preparation.

Limestone handling and storage can be carried out with essentially the same equipment conventionally utilized for coal. Commercially available components have been combined in the various demonstration systems, all of which have operated successfully.

In a typical limestone preparation system for utility application, it is assumed that the raw limestone is delivered crushed to the size required by the pulverizers. If this size requirement cannot be met by the supplier, a crusher will be required at the front end of the system. The crushed limestone is delivered via a vibrating feeder to a bucket elevator which dumps the material into the elevated raw limestone storage bin. The limestone is delivered by gravity through a rotary feeder to the pulverizer(s).

Pulverizers are designed to provide a selected particle size when grinding materials of a similar hardness and density. Because limestone is denser than most coals, simultaneous pulverization of limestone and coal would

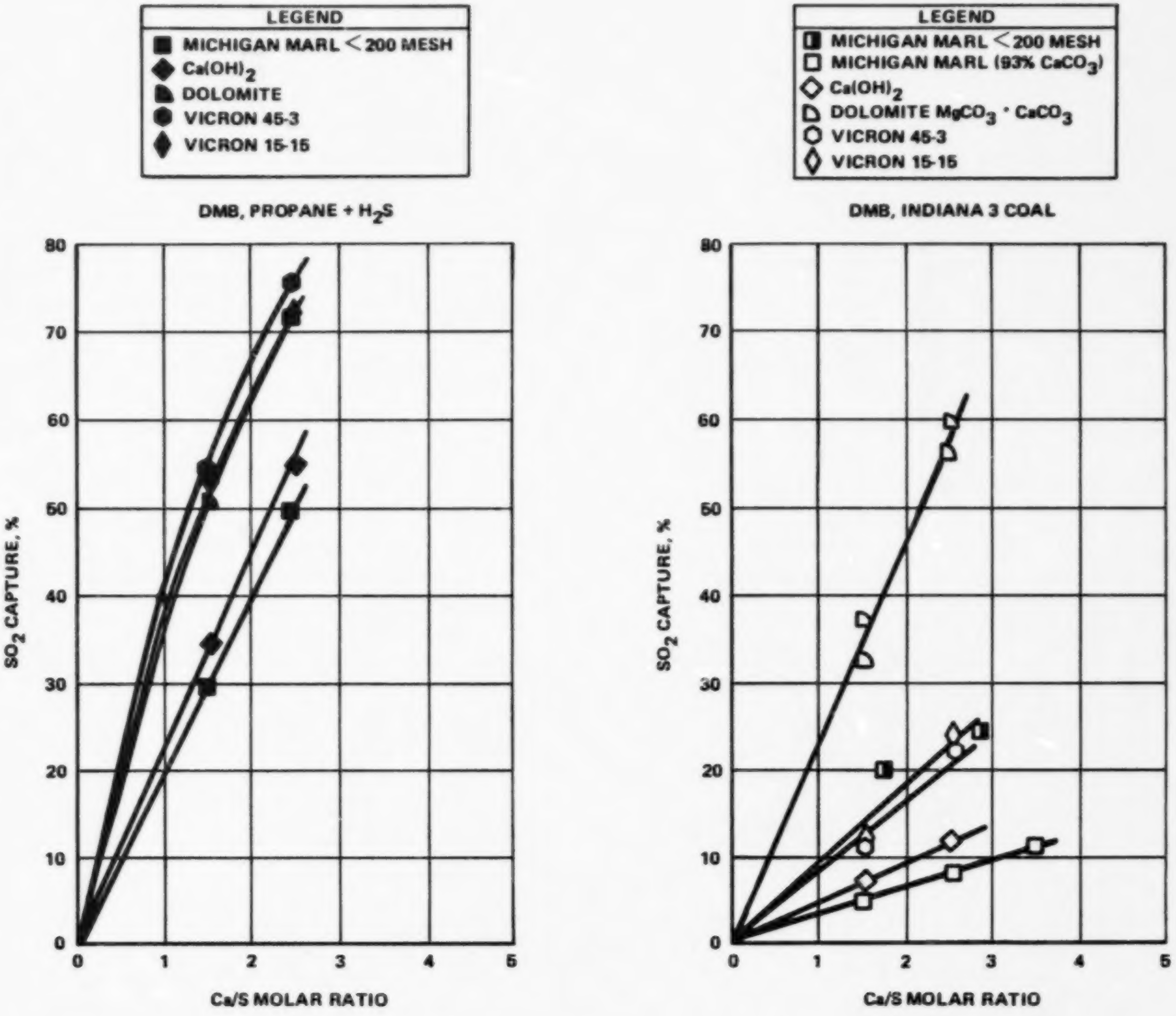


Figure 27. Comparison of SO_2 absorption characteristics with gaseous and coal fuels (sorbent with fuel, uncooled) (Ref. 17).

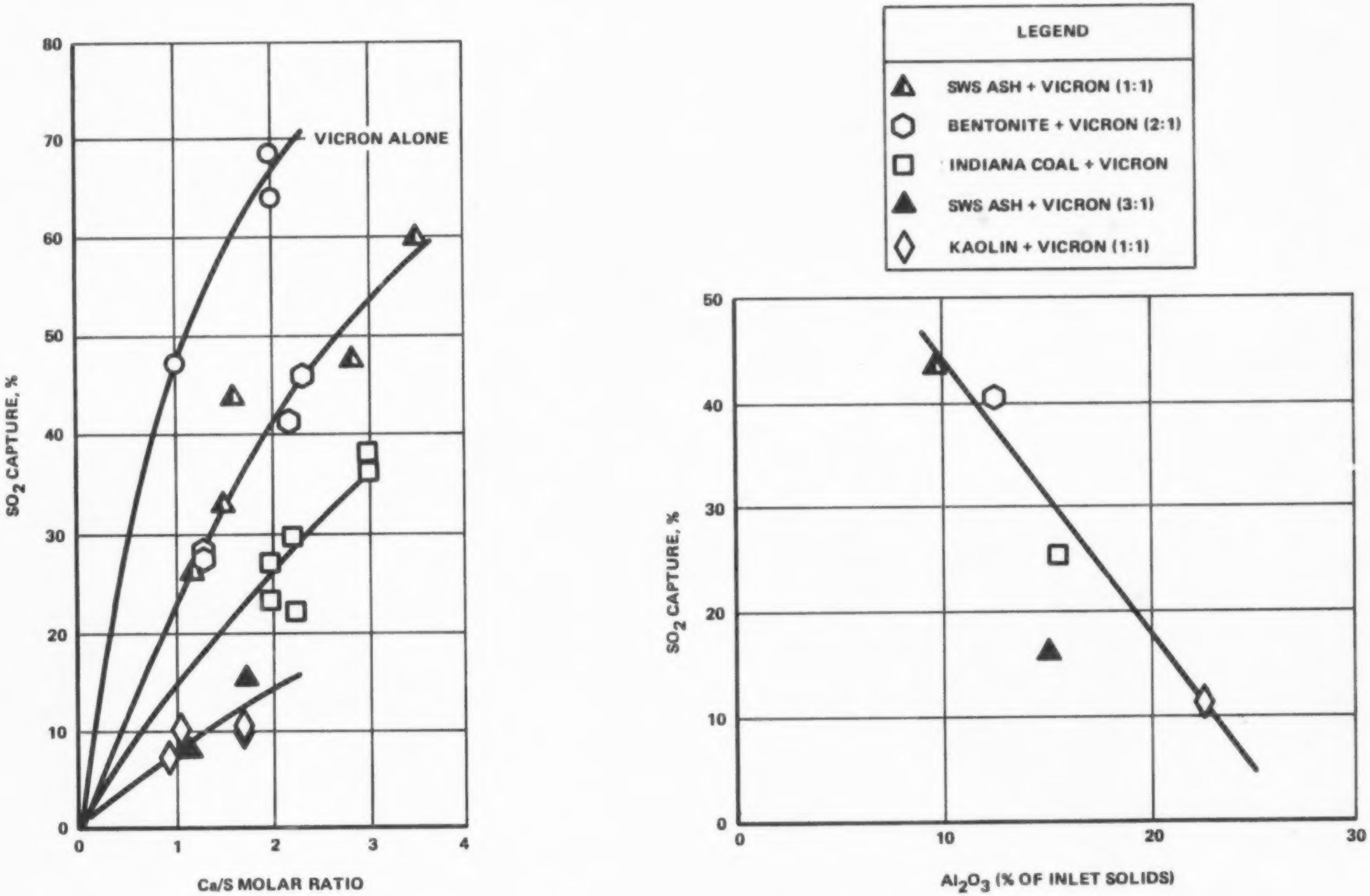


Figure 28. Effect of Al_2O_3 content in ash on SO_2 capture (Ref. 17).

result in excessive mill rejects if adequate control on limestone particle size was maintained. Therefore, the limestone preparation system is designed to include separate mills dedicated to limestone pulverization alone.

As discussed above, limestone particle size is a key variable in the overall SO_2 absorption process. Recent studies have shown that maximum reactivity is realized when the grind is such that 90% to 100% passes through a 325 mesh screen. This results in an average particle size less than 20 microns. For this study, the nominal particle size has been set at from 10 to 20 microns. Particle sizes in this range can be achieved with commercially available pulverizers.

Pulverizers can generally be grouped into three categories, each with its advantages and limitations: low speed (ball tube); high speed (impact); and medium speed (roller, bowl). An example of a low speed pulverizer is a ball tube mill. In this type of device, pulverization is caused by thousands of impacts from steel balls inside a slowly rotating drum. An extremely fine powder can be produced, but energy requirements are very high.

An impact mill is an example of a high speed pulverizer. Impact and abrasive action between the material being ground and rotating beater plates effects the pulverization. This type of mill is typically used for high moisture low rank materials, such as peat, which do not require fine particle sizes. Its high maintenance requirements make it unsuitable for high rank coals.

Medium speed mills represent the majority of pulverization equipment. Typical of this group are the C-E Raymond roller and VR mills. In these devices, pulverization is caused by crushing and particle attrition as driven roller assemblies move over stationary rings. These mills can produce a fine product with very low energy requirements. They are best used for indirect feed applications using storage systems, such as in a LIMB application.

In a recently completed LIMB study for EPRI, Ref. (30), C-E recommended roller mills for low-sulfur coals because of their fineness capability (99% through 325 mesh) and VR mills for high-sulfur coals because of their large capacity (typically 75 to 100 tons per hours, versus 5 to 20 tons per hour for roller mills). The maximum attainable fineness for VR mills, however, is only 90% through 200 mesh, which will not produce the required 10 to 20 micron nominal fineness.

From the mills the pulverized limestone may be sent to storage silos, from which it can be transported as required to the boiler, or it may be directly fed to the boiler. The latter option was selected for this study.

The method of transporting the limestone to the boiler is another important consideration which can affect the process economics and sulfur removal efficiency. The two leading alternative methods are dense-phase transport and dilute-phase transport. A dense-phase system delivers about 20 lbs solid/lb air. The advantage of this system is the small amount of air required to transport the sorbent. The major disadvantages are the high pressure loss (roughly 40 to 60 psi) and associated sealing problems. A dilute-phase system, which was selected for this study, requires about 1.5 lbs of air to deliver one pound of sorbent. The pressure drop is in the range of 15 to 30 inches water. The total amount of air required to deliver the sorbent is typically in the range of 1% to 4% of the total air flow required. The split between primary air, overfire air, and sorbent transport air will have to be carefully controlled to maximize combustion and SO_2 removal efficiencies.

SECTION 5

EFFECTS OF SORBENT INJECTION ON BOILER OPERATION

5.1 Slagging and Fouling

A major unanswered question regarding furnace limestone injection is its impact on ash deposition in the radiant section (slagging) and convective passes (fouling) of a steam generator. Furnace limestone injection increases the mass and changes the chemical composition of the ash, and therefore influences the character of ash deposits in various sections of a steam generator. The distribution and quantity of these deposits and their chemical and physical properties have an important impact on heat transfer performance and steam generator design. These factors, therefore, have a major bearing on the feasibility and cost of furnace limestone injection. An investigation was conducted to determine the effects of limestone injection under low-NO_x firing conditions on the slagging and fouling behavior of two representative coals, Ref. (19).

The tests were conducted in the Fireside Performance Test Facility (FPTF) at Combustion Engineering, Inc's Kreisinger Development Laboratory. As a preliminary to the long duration fireside performance tests, short term tests were conducted to evaluate the effects of furnace operating variables on NO_x and SO_2 emissions reduction to optimize furnace operation for the fireside performance investigations. The emissions reduction tests and fireside performance tests are both reviewed in this section.

5.1.1 Description of the Tests

Figure 29 is a schematic of the test facility. The furnace is bottom-fired with three elevations of overfire air injectors, providing staged combustion capabilities. Limestone may be added through one of three furnace locations. Located in the radiant section of the furnace is a simulated waterwall test surface which is used to study lower furnace slagging behavior. In the convective section, four banks of air-cooled probes are used to

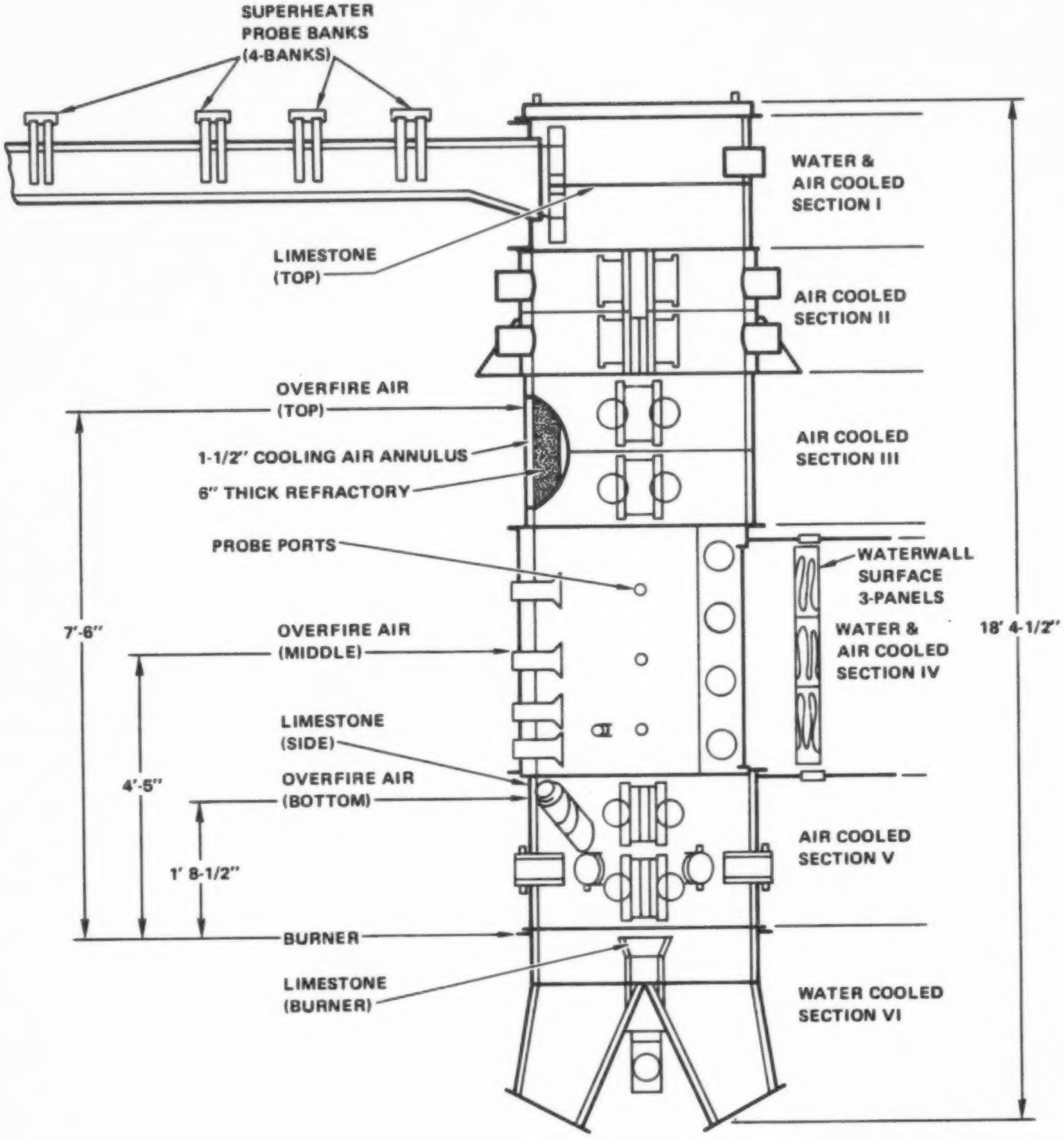

SUPERHEATER
PROBE BANKS
(4-BANKS)
WATER &
AIR COOLED
SECTION I
LIMESTONE
(TOP)
AIR COOLED
SECTION II
OVERFIRE AIR
(TOP)
1-1/2" COOLING AIR ANNULUS
6" THICK REFRACTORY
AIR COOLED
SECTION III
PROBE PORTS
WATERWALL
SURFACE
3-PANELS
WATER &
AIR COOLED
SECTION IV
7'-6"
OVERFIRE AIR
(MIDDLE)
18' 4-1/2"
LIMESTONE
(SIDE)
4'-5"
OVERFIRE AIR
(BOTTOM)
AIR COOLED
SECTION V
1' 8-1/2"
BURNER
LIMESTONE
(BURNER)
WATER COOLED
SECTION VI

Figure 29. Fireside performance test furnace.

simulate boiler superheater tubes and evaluate convective section fouling behavior. The furnace has a maximum firing capacity of approximately 4.0 x 10^6 Btu/hr which corresponds to roughly 350 lb/hr of pulverized coal feed. The test furnace is designed to simulate commercial boiler time/temperature history including flame temperatures from 1900°F to 3000°F and residence times from 1 to 2.5 seconds. Gas samples are obtained at the outlet of the convective pass in a gas temperature region of approximately 1000°F using a sampling probe designed to separate gas from fly ash particles. The gas sample is quenched rapidly to approximately 200°F to prevent further chemical reaction and avoid contamination by water vapor which can absorb SO_2 present in the sample. The gas analysis system measures concentrations of NO_x, SO_2, CO_2, CO and O_2.

Slagging behavior in the lower furnace is evaluated by measurements and observations conducted on the FPTF waterwall panel simulating waterwall heat transfer surface. Deposit cleanability is determined by an on-line sootblower employing 150 psig air as the cleaning medium. The heat flux after sootblowing and its comparison to clean panel heat flux are indications of cleaning effectiveness. Observations are made to identify the physical characteristics of the deposits. Deposits which are dry or lightly sintered are, in general, friable and amenable to sootblowing. Highly sintered or molten deposits are generally more difficult to control by conventional sootblowing.

The fouling behavior is evaluated by measurements and observations in the FPTF convective section. Deposit cleanability is determined by on-line measurements of deposit-to-tube bonding strength (force meter), which gives bonding strength measurements on a relative basis. Deposit accumulation rate is determined in two manners: by the frequency of sootblowing and by weighing the quantity of deposits accumulated in an eight hour period. Physical characteristics of the deposits are observed. Friable deposits are more readily removed than sintered deposits.

Tests were conducted on two coals. The first was a moderate ash (11.2%), high sulfur (3.4%) high iron (14.0% Fe_2O_3 in ash) Midwestern Illinois #6 high volatile C bituminous coal (Baldwin). The second was a low ash (4.9%), low sulfur (0.5%) high sodium (6.9% Na_2O in ash) Western subbituminous coal (WSUB) from Montana.

5.1.2 NO_x and SO_2 Emission Results

With conventional (unstaged) firing conditions during both the Baldwin coal and WSUB coal tests, NO_x values consistently fell in the range of 1000 to 2000 ppm (3% O_2, dry basis). Through staged combustion, the NO_x concentration during the Baldwin coal tests was reduced to as low as 285 ppm and during the WSUB coal tests to as low as 146 ppm. Primary stage stoichiometry and residence time were dominant factors in limiting NO_x.

SO_2 removal was a function of Ca/S molar ratio, residence time, and operating temperatures. The latter two effects were combined to calculate the residence time in the SO_2 removal "window" between 2300°F and 1800°F. The effects of Ca/S molar ratio and residence time are shown for Baldwin and WSUB coals in Figures 30, 31 and 32. The burner zone stoichiometries shown in the figures were the optimums selected to minimize both NO_x and SO_2 emission. At a specified Ca/S molar ratio, based on calcium added with the limestone, SO_2 removal was higher for the WSUB coal. This is because the calcium in the ash and the alkali present in the furnace deposits provide additional SO_2 removal capability with the WSUB coal.

5.1.3 Slagging and Fouling Experience

During conventional firing of Baldwin coal, slagging behavior in field units and at the FPTF is generally satisfactory but there is little margin to operate beyond the design load. Operating the FPTF in the low-NO_x firing mode without limestone injection increased the severity of slagging due to reducing conditions in the firing zone and the resultant decrease in the fusion temperature of the iron-rich ash. Limestone injection had a favorable effect on slagging behavior by increasing the fusion temperature of the ash. During operation of the low-NO_x burner with limestone injection, deposit build-up on

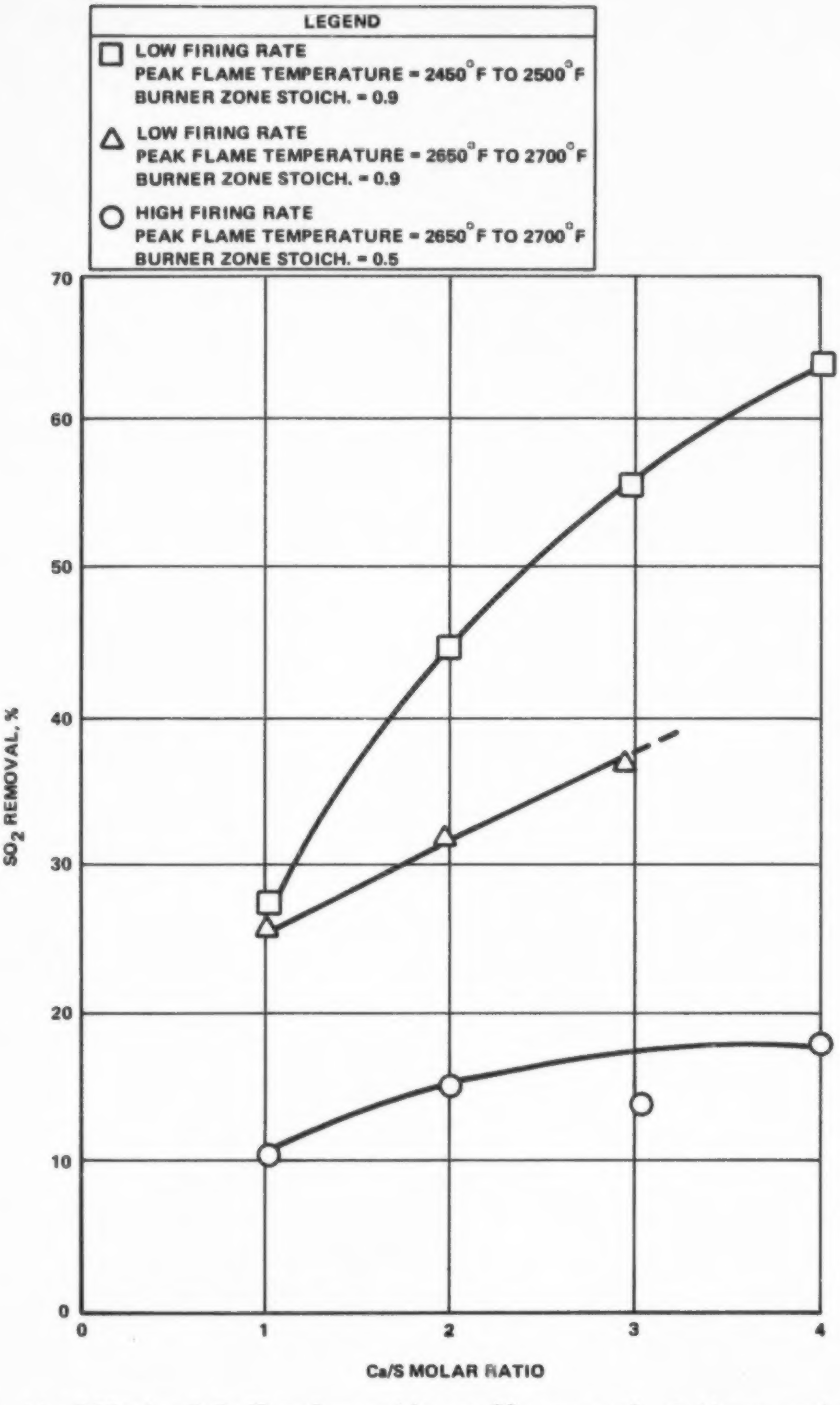


Figure 30. Effect of Ca/S molar ratio on SO_2 removal at three combustion conditions firing Baldwin coal.

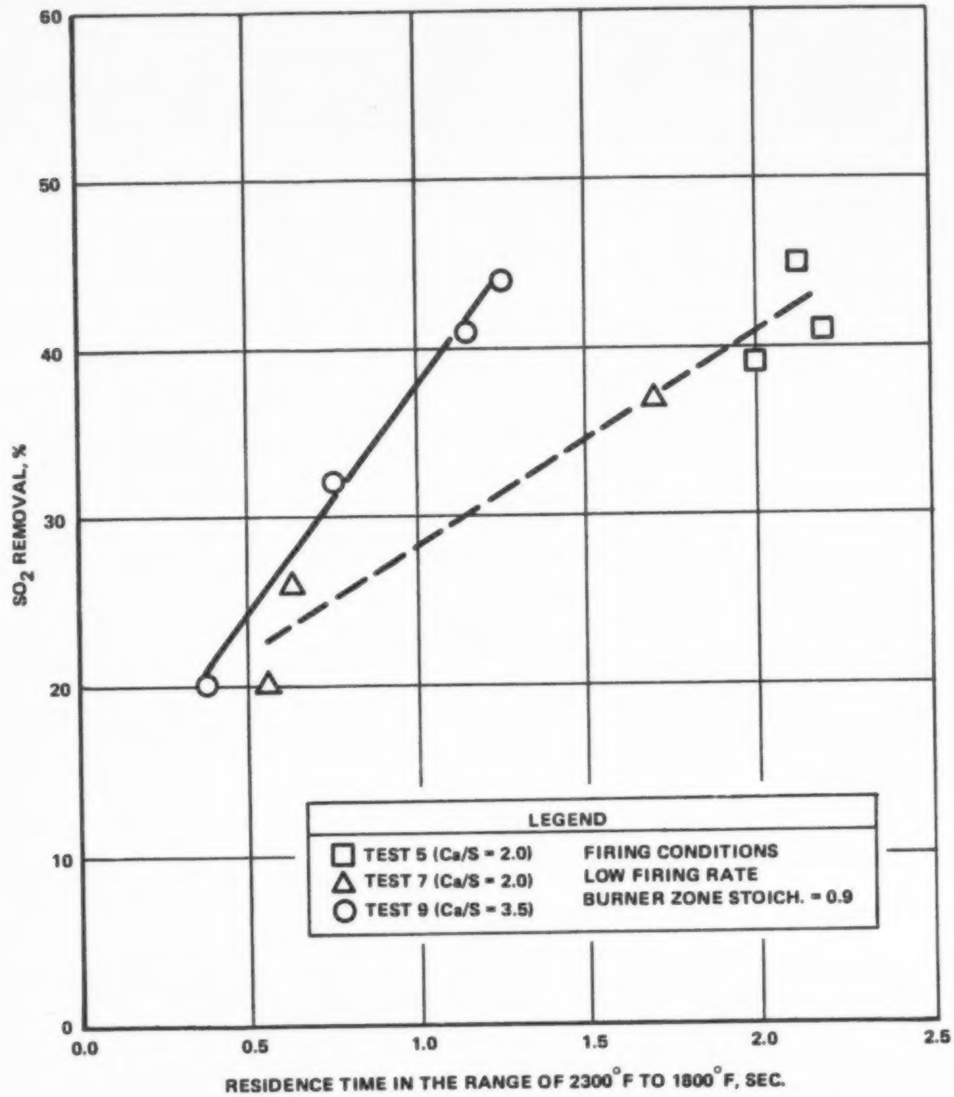


Figure 31. Percent of SO_2 removal vs. residence time in the temperature range of 2300°F to 1800°F during Baldwin coal testing.

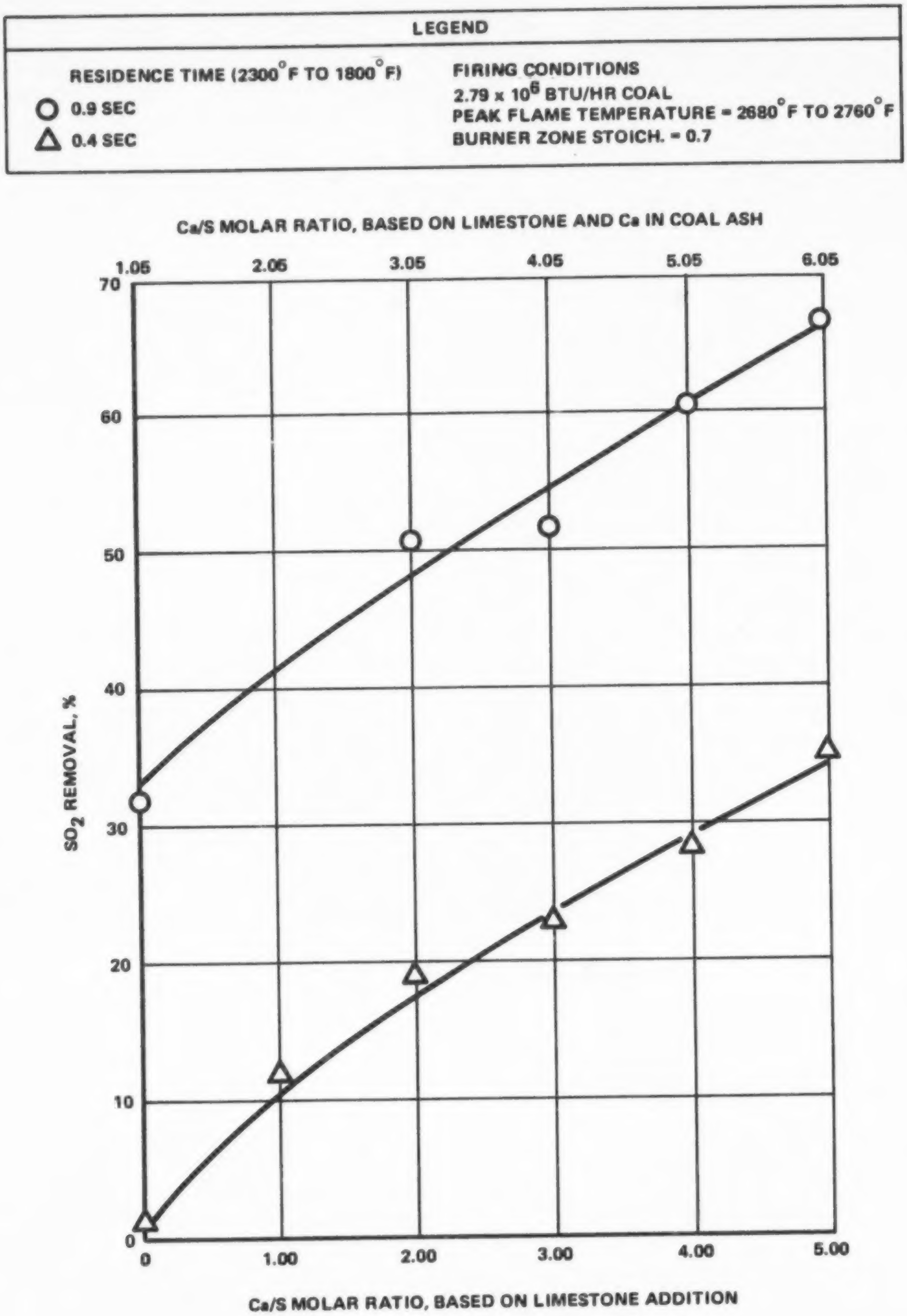


Figure 32. Effect of Ca/S molar ratio on SO_2 removal at two residence times in the temperature range of 2300°F to 1800°F while test firing a western subbituminous coal.

the waterwall simulation panel was more rapid but the deposit build-up was controllable by more frequent sootblowing. Convective section probe deposit build-up rates increased during low-NO_x firing with Baldwin coal. Deposit friability increased and deposit-to-tube bonding strength decreased as limestone addition was increased. During limestone injection, deposit build-up in the convective section was controllable by sootblowing. Conventional firing of WSUB coal results in thin molten deposits which require frequent sootblowing. When WSUB coal was fired in the low-NO_x firing configuration without limestone injection, molten waterwall deposits were formed which were marginally removable by sootblowing. With limestone injection and the low-NO_x firing system the waterwall deposits were more friable. The deposit build-up was more rapid, but the deposit buildups were controllable by sootblowing. Firing WSUB coal in the low-NO_x mode without limestone injection resulted in hard, sintered deposits in the convective section which were not controllable by sootblowing. While firing WSUB coal in the low-NO_x mode with limestone addition, the deposit build-up increased but the deposits were controllable by sootblowing.

5.1.4 Effect on Ash Fusion Temperature

Ash fusion temperature (AFT) tests were conducted at Combustion Engineering for several types of coal with varying Ca/S molar ratios. The tests were run to show the effects of sorbent injection on ash performance as measured by AFT. An increase in AFT indicates a decrease in slagging and fouling tendency. The results indicate that the AFT behavior varies with both coal types and Ca/S molar ratio. This is shown in Figures 33 through 35 for Eastern bituminous, Midwestern bituminous, and Wyoming subbituminous C coals. For Ca/S molar ratios between 0.5:1 and 2:1, the Eastern bituminous coal shows a decrease in AFT while both the Midwestern bituminous and Wyoming subbituminous coals show an increase in AFT. For Ca/S ratios between 2:1 and 4:1, in general, the AFT tends to level off for all three coals.

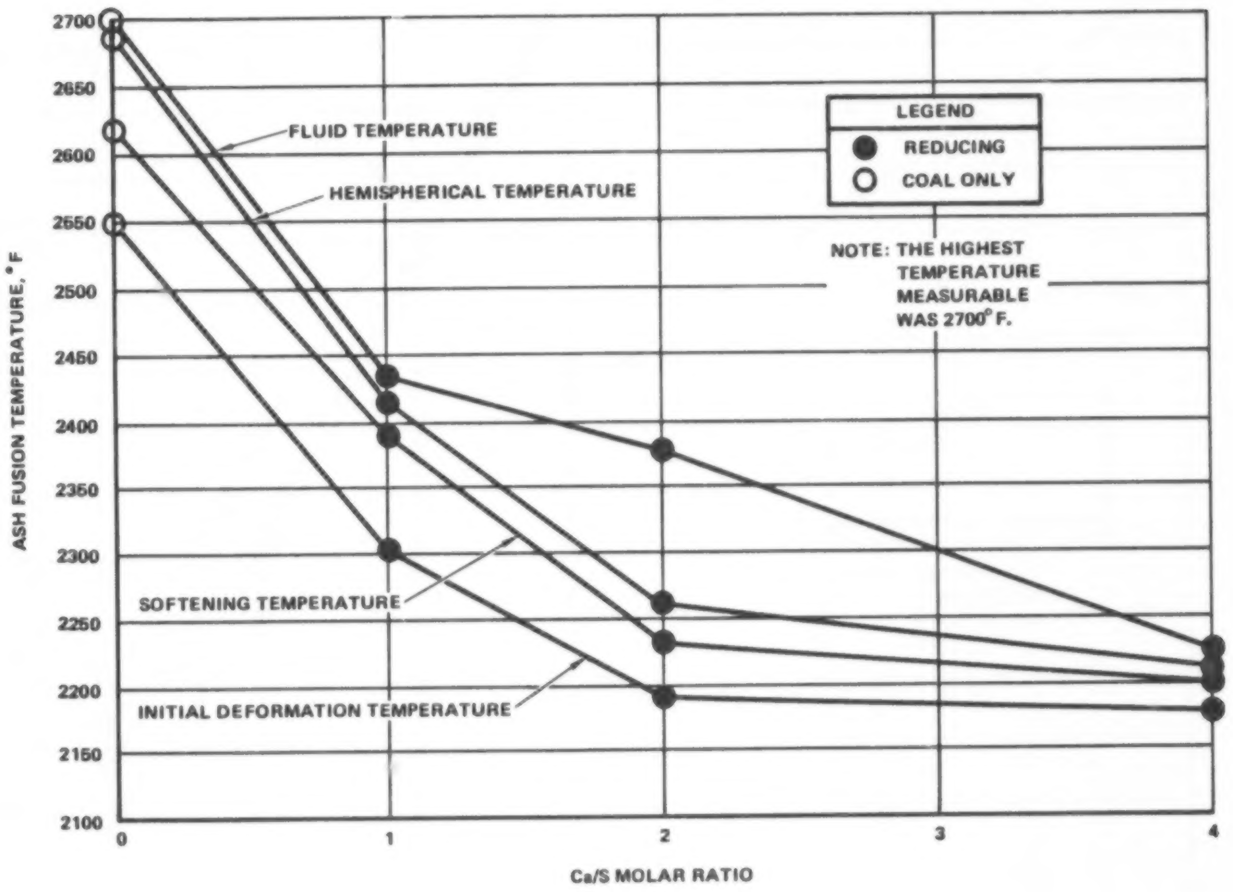


Figure 33. Results of laboratory ash fusion tests with limestone for Eastern bituminous coal.

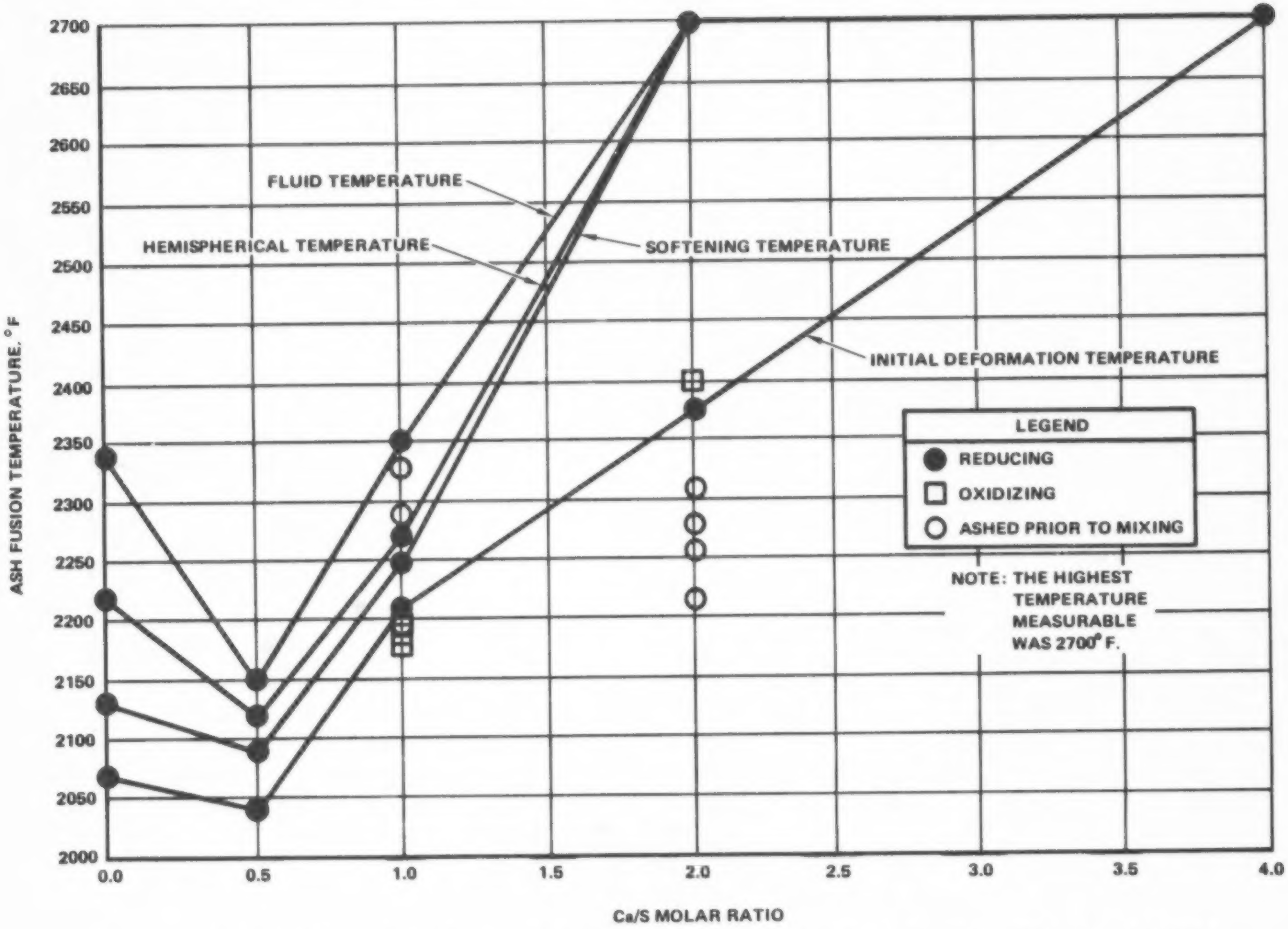


Figure 34. Results of laboratory ash fusion tests with limestone for Midwestern bituminous coal.

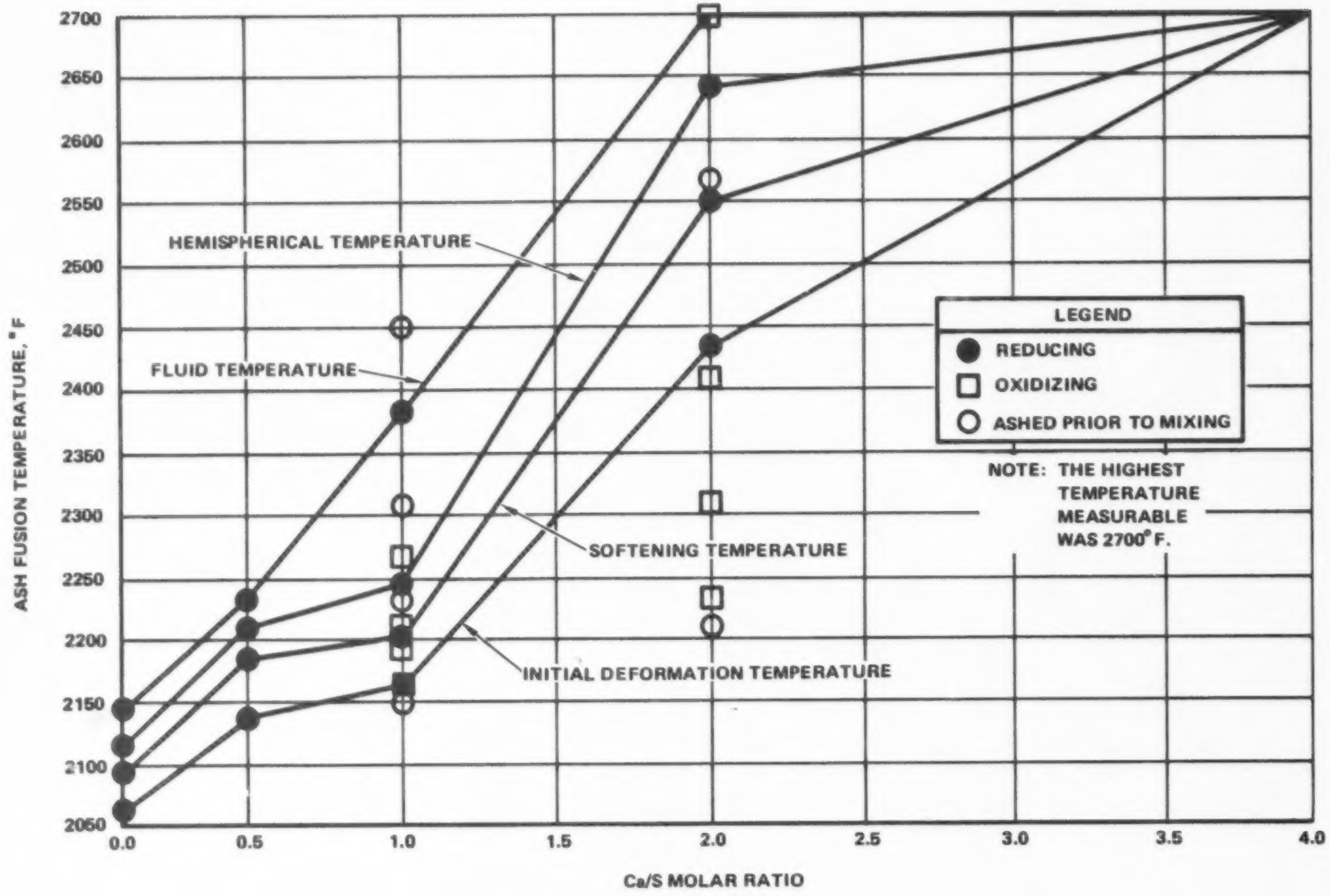


Figure 35. Results of laboratory ash fusion tests with limestone for Wyoming subbituminous C coal.

5.2 Steam Generator Efficiency

The fuel input rate to a steam generator is defined by the efficiency of the steam generator. The efficiency may be defined as either including or not including losses associated with steam generator auxiliaries. For the purpose of this discussion, the boundaries for defining the steam generator efficiency are selected to encompass the heat transfer components. The efficiency can then be defined as the ratio of the heat recovered by the working fluids in the boiler, superheater, reheater, economizer, and air heater, to the total heat input. It may also be defined as the total heat input minus the various losses, expressed as percentages. Expressing this mathematically, the boiler efficiency is then Eff $\% = 100\% - \Sigma$losses, %. The following losses are normally considered in the calculations:

1. Dry gas loss - This quantity represents the sensible energy of the gaseous products, exclusive of any resulting moisture leaving the steam generator envelope.

2. Air moisture loss - This quantity represents the sensible energy of the water vapor contained in the air entering the steam generator envelope.

3. Fuel moisture loss - This quantity accounts for the evaporation and superheating of the moisture entering in, and produced from, the fuel entering the steam generator envelope.

4. Carbon loss - This quantity represents the chemical energy contained in carbon particles entrained in the slag and ash leaving the steam generator envelope.

5. Radiation loss - This quantity accounts for the radiation loss from the steam generator to the surroundings.

6. Fly ash heat loss - This quantity accounts for the sensible energy of the fly ash leaving the steam generator envelope.

7. Ash pit loss - This quantity accounts for the sensible energy of the ash collected in the ash pit.

8. Unaccounted for and manufacturer's tolerance losses - Self-explanatory.

The largest individual contributor to the boiler losses is the dry gas loss which depends on the flue gas temperature leaving the air preheater and the gas mass flow rate.

Sorbent injection introduces several additional losses:

1. Heat required for calcination
2. Increased dry gas loss
3. Increased fly ash heat loss
4. Increased ash pit loss

The magnitude of the efficiency reduction depends on the sulfur content of the coal and the calcium to sulfur ratio. For high-sulfur coals, these losses will be offset to a degree by an efficiency gain resulting from a decrease in the allowable gas temperature leaving the steam generator envelope. This relates to the fact that boiler efficiency is very sensitive to the gas exit temperature, which is largely dictated by the coal sulfur content. This temperature is normally expressed in terms of a minimum air preheater average cold-end temperature (ACET), which is the lowest recommended value to avoid acid corrosion in the air heater. The ACET is defined as the average of the air heater air inlet and gas outlet temperatures. Typically, this temperature is about 155°F for coals containing up to 1% sulfur and 180°F for coals containing 3% to 5% sulfur. The corrosion potential results from the fact that a small portion of the sulfur in the coal forms sulfur trioxide (SO_3), which reacts with the moisture in the flue gas to form sulfuric acid. Sorbent injection reduces the amount of SO_2/SO_3 in the flue gas, resulting in a reduction in the permissible ACET. This temperature reduction results in an increase in boiler efficiency which compensates for the reduction in efficiency due to sorbent injection. These effects should be investigated further.

5.3 Effects of Low-NO_x Firing System on Furnace Performance

Combustion Engineering, Inc is currently marketing three types of low-NO_x firing systems:

- Tangential firing combined with overfire air
- Low-NO_x Concentric Firing System (LNCFS)
- MHI PM Firing System

The application of each system depends on unit design, NO_x level requirements, and economics. Only the LNCFS will be discussed for this study because of its retrofitability and relatively low cost when applied to tangentially fired units.

The LNCFS utilizes the rich fireball concept. It directs the fuel and part of the secondary air into the center of the furnace, with the remaining secondary air injected along the furnace walls. This arrangement produces a fuel-rich combustion zone in the center and increases the residence time of the combustion products before they come into contact with the main body of combustion air surrounding them. In addition to lowering NO_x emissions, the LNCFS produces a thermal environment which should enhance sorbent reactivity for higher sulfur removal. Tests have shown that the LNCFS has no adverse effect on furnace performance. In fact, a reduction in furnace slagging was observed on tested units. This reduction was mainly due to the air sweeping and blanketing the furnace walls.

5.4 Effect of Sorbent Injection on Overall Steam Generator and Component Performance

Sorbent injection greatly increases the quantity and changes the character of the solids passing through a steam generator. As a consequence, because the slagging and fouling characteristics are changed, the overall steam generator efficiency and the performance of the individual components (e.g., superheater and reheater outlet temperatures) may be affected.

It is expected that a low-NO_x firing system plus sorbent injection will result in an improvement in waterwall deposit friability (ease of removal) and sootblower effectiveness for most coals. For retrofit units firing coals which exhibit severe slagging in the presence of sorbent, the steam generator capacity may be adversely affected. If slagging is excessive, it may inhibit waterwall absorption, reducing evaporation and raising the furnace outlet gas temperature (FOT, defined as the temperature at the bottom of the vertical radiant surfaces (panels) suspended in the upper section of the furnace). If increased sootblowing frequency is not adequate to alleviate this problem, the unit capacity may have to be reduced. For new units, these potential waterwall slagging problems may be minimized by including larger furnaces in the design.

Superheater and reheater outlet temperatures are determined by FOT, gas flow, and the degree of fouling on the heat transfer surfaces. An FOT which is higher than the design value results in higher superheater and reheater temperatures. These temperatures are normally controlled by manipulating burner tilts, excess air, and superheater/reheater spray. However, since the latter two options have considerable influence upon the overall plant efficiency, their use should be minimized. As in the lower furnace section, the increase in solids loading with sorbent injection will result in an increase in the deposit rate on superheater and reheater surfaces. This will require an increase in the sootblowing frequency. For new units, design considerations may include the use of wider spacing in the convective section to prevent tube bridging from deposit buildup and/or the addition of widely spaced superheater panels to reduce the gas temperature before the gas enters the convective sections.

Sorbent injection will result in an increase in the gas/solids mass flow through the steam generator. In retrofit applications, this will result in an increase in the mass velocity through the unit. While this increased velocity will cause an increase in the convective heat transfer coefficient, it is important to keep the velocity within acceptable limits to minimize tube erosion. The maximum allowable gas velocity depends on the erosion characteristics of the flyash and the quantity of flyash. For conventional coal-fired units, the general practice is to limit the maximum velocity to approximately 50 to 70 feet per second. With sorbent injection, this velocity may have to be reduced, due to the increased solids loading and the ash erosion characteristics (not yet known). In a retrofit application, widening the spacing between tube elements to lower gas velocity may not be economical. A reduction in unit capacity may therefore be required to keep the gas velocity within acceptable limits. For new units, the convective section can be designed with wider transverse spacing to keep the gas velocity below the maximum value.

The final heat transfer sections located within the steam generator envelope are the economizer and the air preheater. Prior to the early 1970's, most economizers were of a finned-tube staggered arrangement design. The significant increase in solids loading with sorbent injection would make this type of economizer very susceptible to plugging. Therefore, for retrofit units which incorporate finned economizers, it may be necessary to replace the economizer with one featuring bare tubes with an in-line arrangement. This will generally result in a much larger economizer to accomplish the same amount of heat transfer. In some cases, installation of a significantly larger economizer may be impractical due to space limitations. An alternative in these cases would be to use continuously finned tubes with an in-line arrangement. While some additional space may be required, it will be substantially less than with a bare tube economizer. Tests have shown that with sufficient sootblowing capability, plugging should not be a serious problem. Should neither of these alternatives be feasible due to severe space limitations, the retrofit economizer will be undersized. Insufficient economizer surface results in a reduction in the quantity of steam evaporated in the steam generator. For some coals, this may be fully or partially overcome by tilting the coal nozzles downward and/or increasing the frequency of sootblowing in the furnace to remove accumulated deposits and increase waterwall absorption.

Studies performed on air heaters during limestone or dolomite injection have shown that, with reduced SO_3 levels in the gas stream, plugging, as indicated by pressure drop across the air heater, is reduced to a minimum. While the ash loading is increased, the actual buildup decreases, and that which forms tends to be fluffy and readily removable. Low temperature probe studies conducted during full-scale limestone furnace injection tests showed that SO_3 was reduced to a level sufficiently low to eliminate corrosion and produce deposits which are dry and generally non-adherent.

For units which will require increased sootblowing frequency and/or sootblower coverage, it may be necessary to upgrade the capacity of the sootblowing system. It should be noted, however, that the additional steam of air consumption required to accomplish this will have a negative impact upon steam generator and overall plant efficiency.

The increased gas flow rate through the convective sections which is associated with sorbent injection may result in higher draft losses for retrofit units. This must be considered in evaluating the capacity of existing induced draft fans.

SECTION 6

PILOT-PLANT AND FULL-SCALE LIMB INVESTIGATIONS

6.1 MHI Low-NO_x PM Burner

The PM (pollution minimum) burner was developed by Mitsubishi Heavy Industries, Ltd. (MHI) to reduce NO_x emissions from coal-burning tangentially fired utility steam generators. The primary mechanism by which NO_x reduction is achieved in the PM burner is staged admission of air which results in a lower temperature, fuel-rich environment in the immediate vicinity of the coal nozzles. The burner was tested by MHI in a 4 ton/hour burner test facility.

In the second phase of the PM burner test program, the SO_2 removal effectiveness of furnace limestone injection, in combination with the low-NO_x burner, was evaluated in the 4 ton/hour test facility, Ref. (18). This testing was supported by basic SO_2 removal investigations in a heated tube reactor. The heated tube reactor studies are discussed in Section 4. The investigation of SO_2 removal in the 4 ton/hour low-NO_x furnace is discussed in this section.

6.1.1 Description of Tests

The furnace used in these tests, shown in Figure 36, is a water-cooled cylindrical vessel 4.4 meters (14.4 feet) in diameter and 20 meters (65.6 feet) long. The inner surface is partially covered with refractory to simulate the heat absorption profile which would exist in a tangentially fired furnace. The burner used in the limestone injection tests is shown in Figure 37. The concentrated coal nozzle located at the middle of the burner is provided with a flame holder at its outlet to stabilize the ignition of the rich coal/air mixture injected from this nozzle. Separating gas recirculation (SGR) nozzles are arranged above and below the concentrated coal nozzle. A weak (or fuel-lean) nozzle, an auxiliary air nozzle, and an oil nozzle are located below the lower SGR nozzle. An auxiliary air nozzle and an oil nozzle are located above the upper SGR nozzle. The upper SGR nozzle is to retard mixing of secondary air from the upper auxiliary air nozzle with the concentrated coal nozzle flame and the lower SGR nozzle is to retard mixing of the weak nozzle's flame with the concentrated coal nozzle flame. Four methods of limestone injection were used as shown in Figure 38. In the Type A limestone injection method, coal and air were carried together by the primary air. In the Type B injection method, the limestone was injected through the auxiliary air compartment above the concentrated coal nozzle. In the overfire air injection method, Type C, the limestone was injected tangentially with up to 20% of the total air flow at a location 12.6 meters (41.3 feet) from the burner face. In the gas recirculation injection method, Type D, the limestone was injected tangentially with up to 15% of the flue gas flow at a location 12.6 meters (41.3 feet) from the burner face. An overall flow schematic showing the location of sampling points is given in Figure 39. Tests were run with a bituminous coal containing 3.1% sulfur (high sulfur) and a subbituminous coal containing 0.65% sulfur (low sulfur).

6.1.2 SO_2 Capture Performance

The tests evaluated the effects of limestone injection location, calcium to sulfur molar ratio (Ca/S), sulfur content of the coal, fineness of the limestone, firing rate, excess oxygen and quantity of overfire air. Figure 40 shows that limestone injected with the overfire air, Type C, gave the best SO_2 capture. This injection method introduces the limestone downstream of the high temperature region at the burner face and provides the highest quantity of oxygen in the region where the reaction of SO_2 with CaO takes place. The remainder of the MHI test program was conducted with the Type C injector. Figures 40 and 41 indicate that SO_2 capture increases with increasing Ca/S ratio but the effect of Ca/S ratio is non-linear at high Ca/S ratio. Sulfur removal rate is higher for the high sulfur coal. Figure 42 shows that there is negligible effect of limestone fineness over the range tested. In the MHI tests the firing rate of 100% gave the best SO_2 removal rate as shown in Figure 43. At the 120% firing rate the residence time was decreased whereas at the 80% firing rate the temperature in the reaction zone was less than optimum. Figures 44 and 45 indicate that SO_2 capture is

89

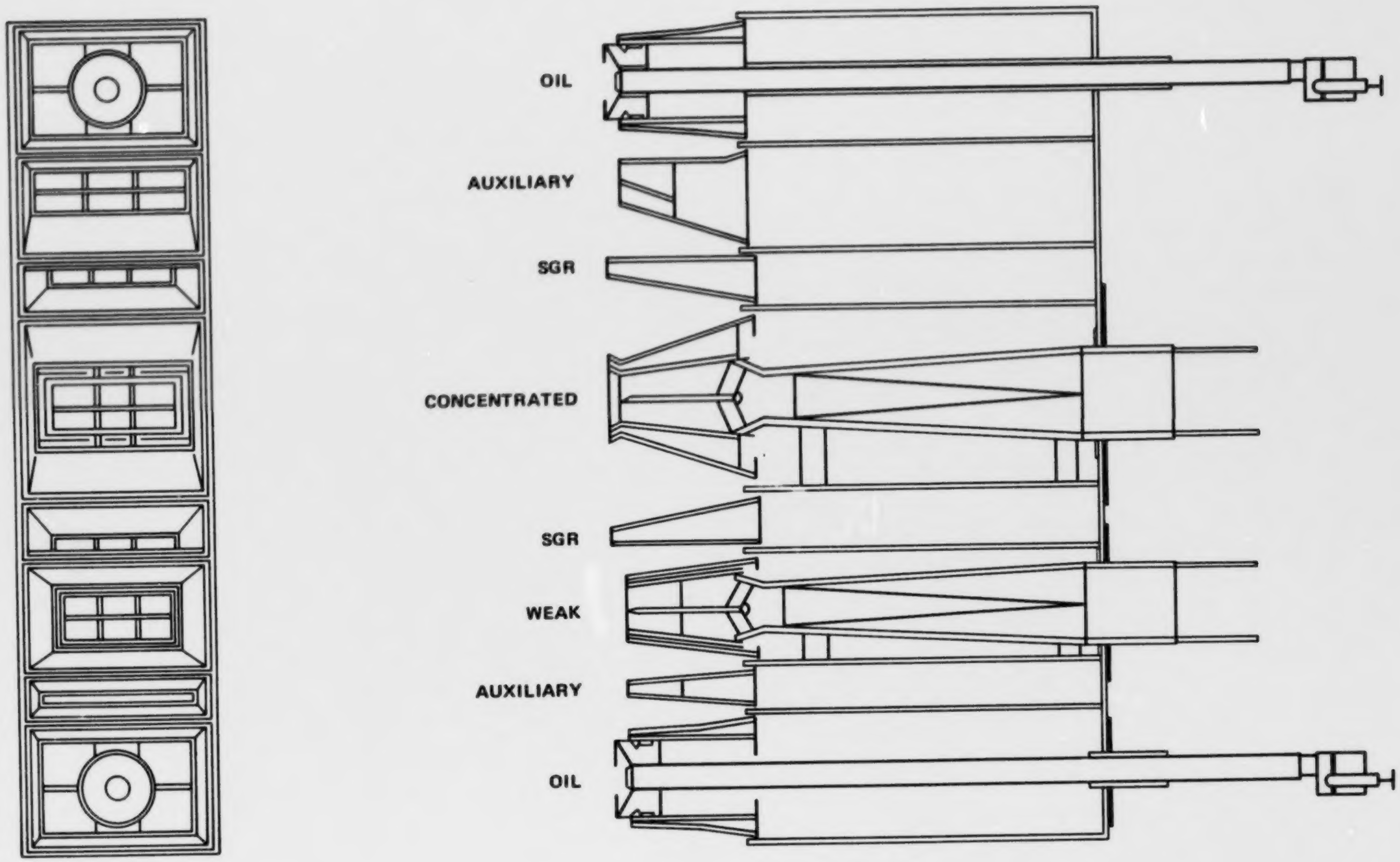

OIL
AUXILIARY
SGR
CONCENTRATED
SGR
WEAK
AUXILIARY
OIL

Figure 37. Configuration of coal-fired low-NO_x PM burner (Ref. 18).

SGR nozzle. The upper SGR nozzle is to retard mixing of secondary air from the upper auxiliary air nozzle with the concentrated coal nozzle flame and the lower SGR nozzle is to retard mixing of the weak nozzle's flame with the concentrated coal nozzle flame. Four methods of limestone injection were used as shown in Figure 38. In the Type A limestone injection method, coal and air were carried together by the primary air. In the Type B injection method, the limestone was injected through the auxiliary air compartment above the concentrated coal nozzle. In the overfire air injection method, Type C, the limestone was injected tangentially with up to 20% of the total air flow at a location 12.6 meters (41.3 feet) from the burner face. In the gas recirculation injection method, Type D, the limestone was injected tangentially with up to 15% of the flue gas flow at a location 12.6 meters (41.3 feet) from the burner face. An overall flow schematic showing the location of sampling points is given in Figure 39. Tests were run with a bituminous coal containing 3.1% sulfur (high sulfur) and a subbituminous coal containing 0.65% sulfur (low sulfur).

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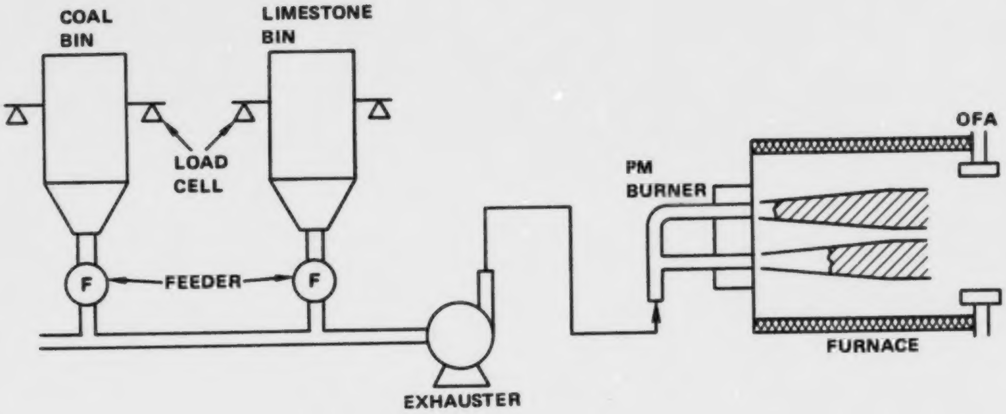
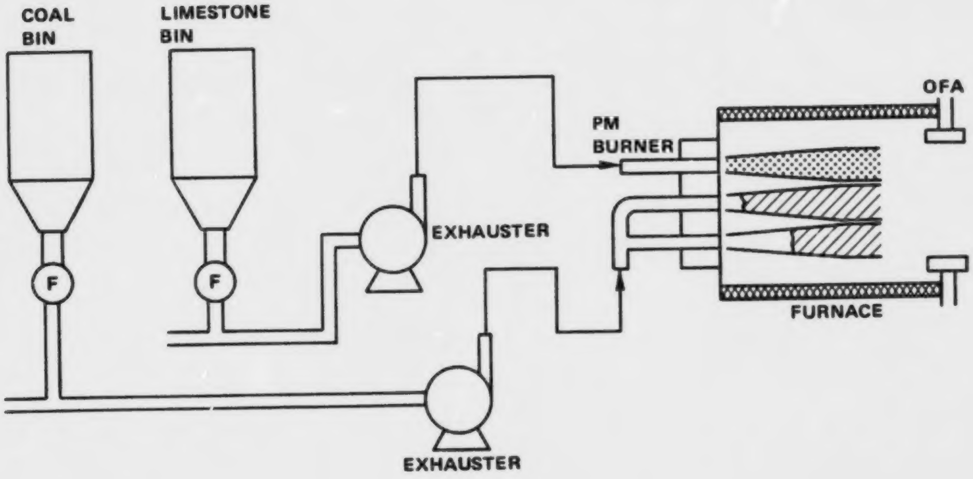
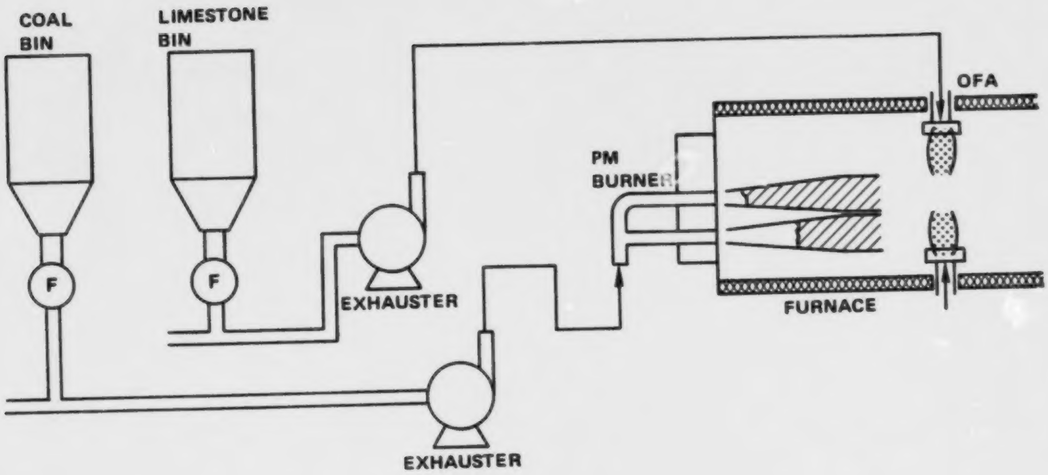

COAL
BIN
LIMESTONE
BIN
LOAD
CELL
F
FEEDER
F
EXHAUSTER
PM
BURNER
OFA
FURNACE

TYPE A: COAL/LIMESTONE MIXING SYSTEM

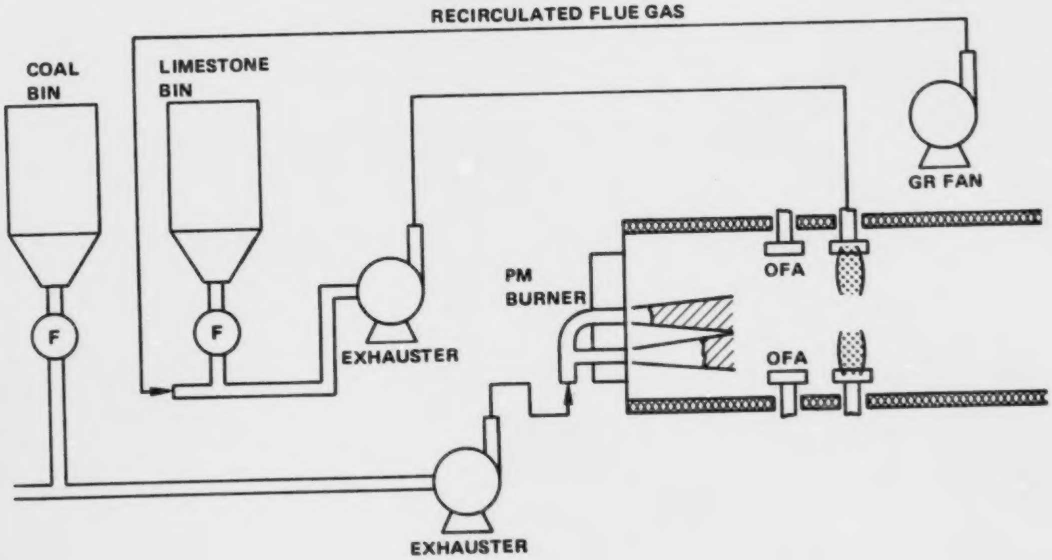

COAL
BIN
LIMESTONE
BIN
F
F
EXHAUSTER
PM
BURNER
OFA
FURNACE
EXHAUSTER

TYPE B: THROUGH THE AUXILIARY AIR NOZZLE SYSTEM

Figure 38. Injection system for coal and limestone (Ref. 18).


COAL BIN
LIMESTONE BIN
F
F
EXHAUSTER
EXHAUSTER
PM BURNER
OFA
FURNACE

TYPE C: THROUGH THE OFA NOZZLE SYSTEM


RECIRCULATED FLUE GAS
COAL BIN
LIMESTONE BIN
F
F
EXHAUSTER
EXHAUSTER
GR FAN
PM BURNER
OFA
OFA

TYPE D: THROUGH THE GR NOZZLE SYSTEM

Figure 38. Injection system for coal and limestone (Ref. 18) (continued).

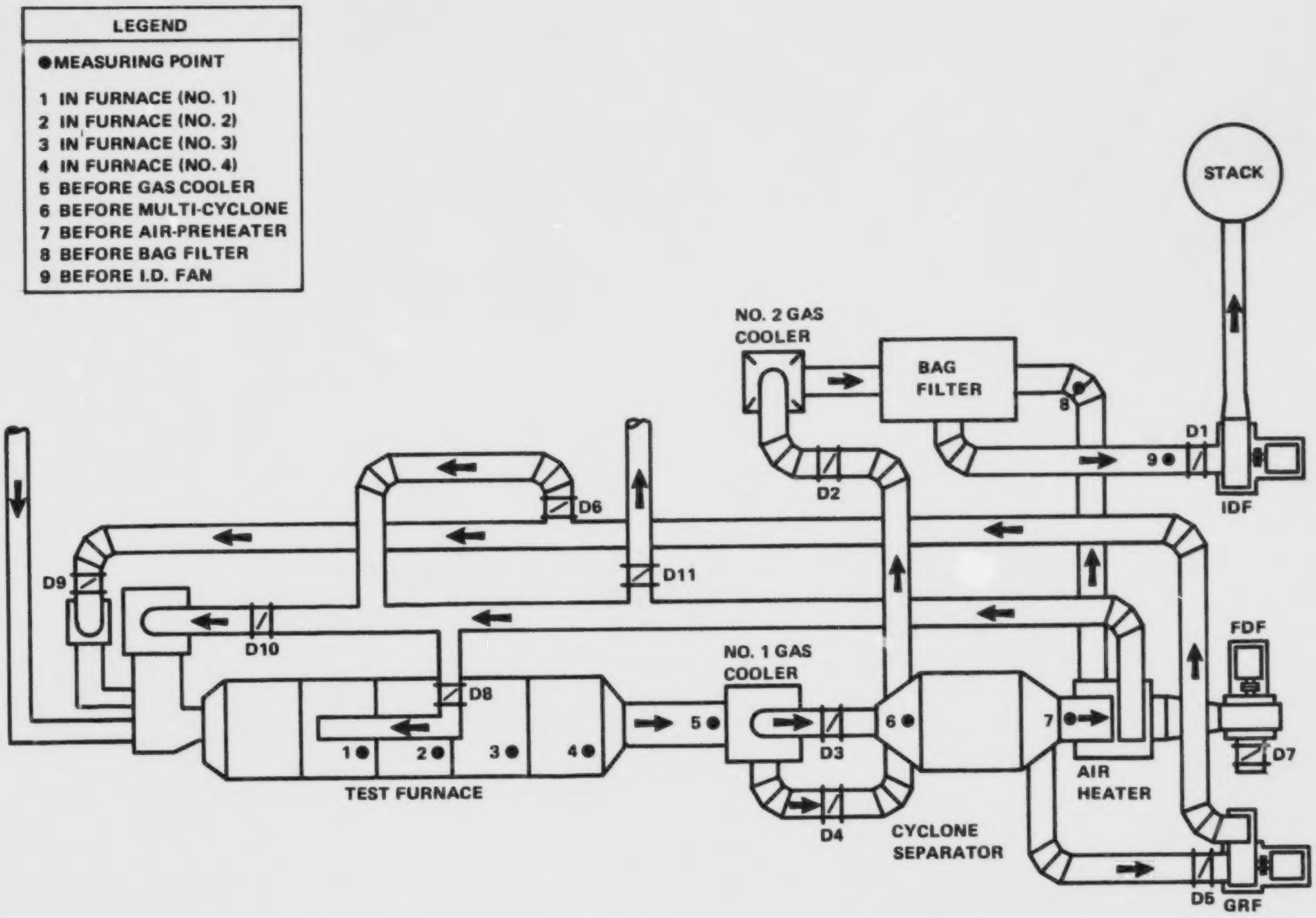

LEGEND
● MEASURING POINT
1 IN FURNACE (NO. 1)
2 IN FURNACE (NO. 2)
3 IN FURNACE (NO. 3)
4 IN FURNACE (NO. 4)
5 BEFORE GAS COOLER
6 BEFORE MULTI-CYCLONE
7 BEFORE AIR-PREHEATER
8 BEFORE BAG FILTER
9 BEFORE I.D. FAN
STACK
NO. 2 GAS COOLER
BAG FILTER
8
D1
9
IDF
D2
D6
D11
D9
D10
FDF
NO. 1 GAS COOLER
D8
5
6
7
1
2
3
4
D3
D7
TEST FURNACE
AIR HEATER
CYCLONE SEPARATOR
D4
D5
GRF

Figure 39. Location of SO_2 measuring points in 4 ton/hr test furnace (Ref. 18).

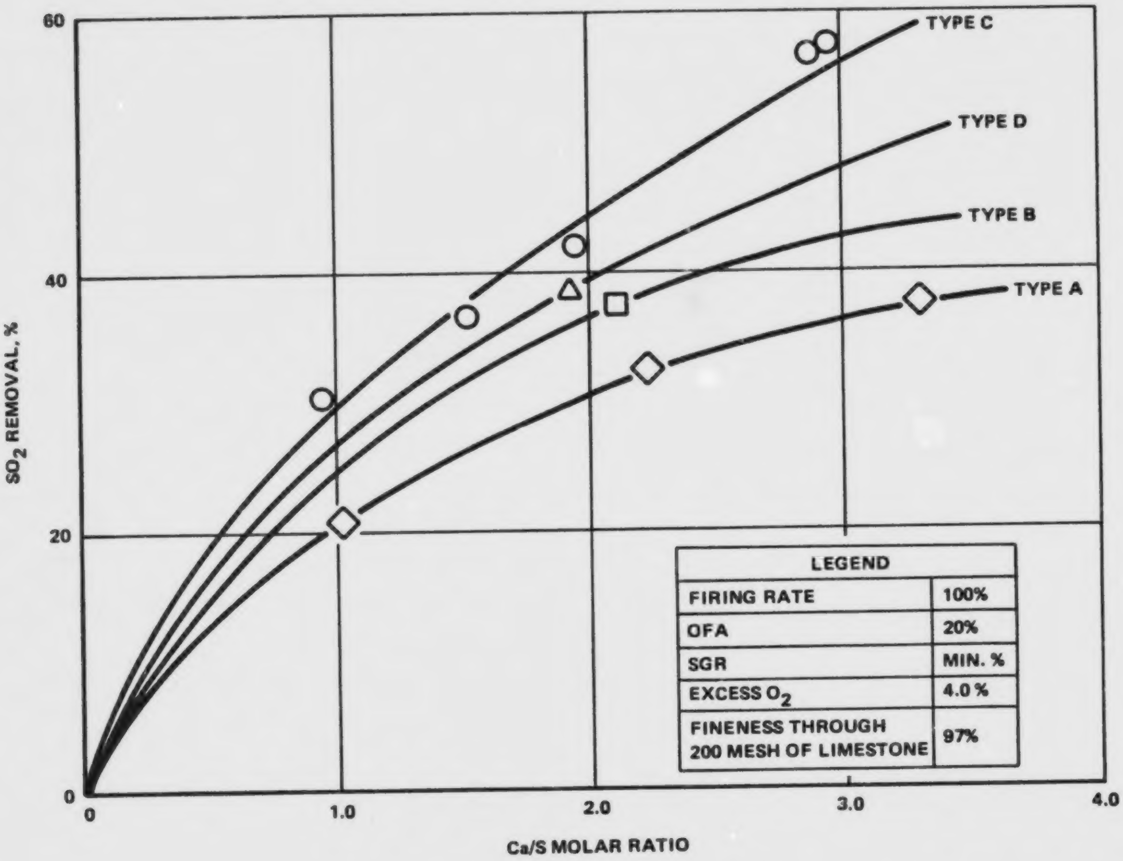
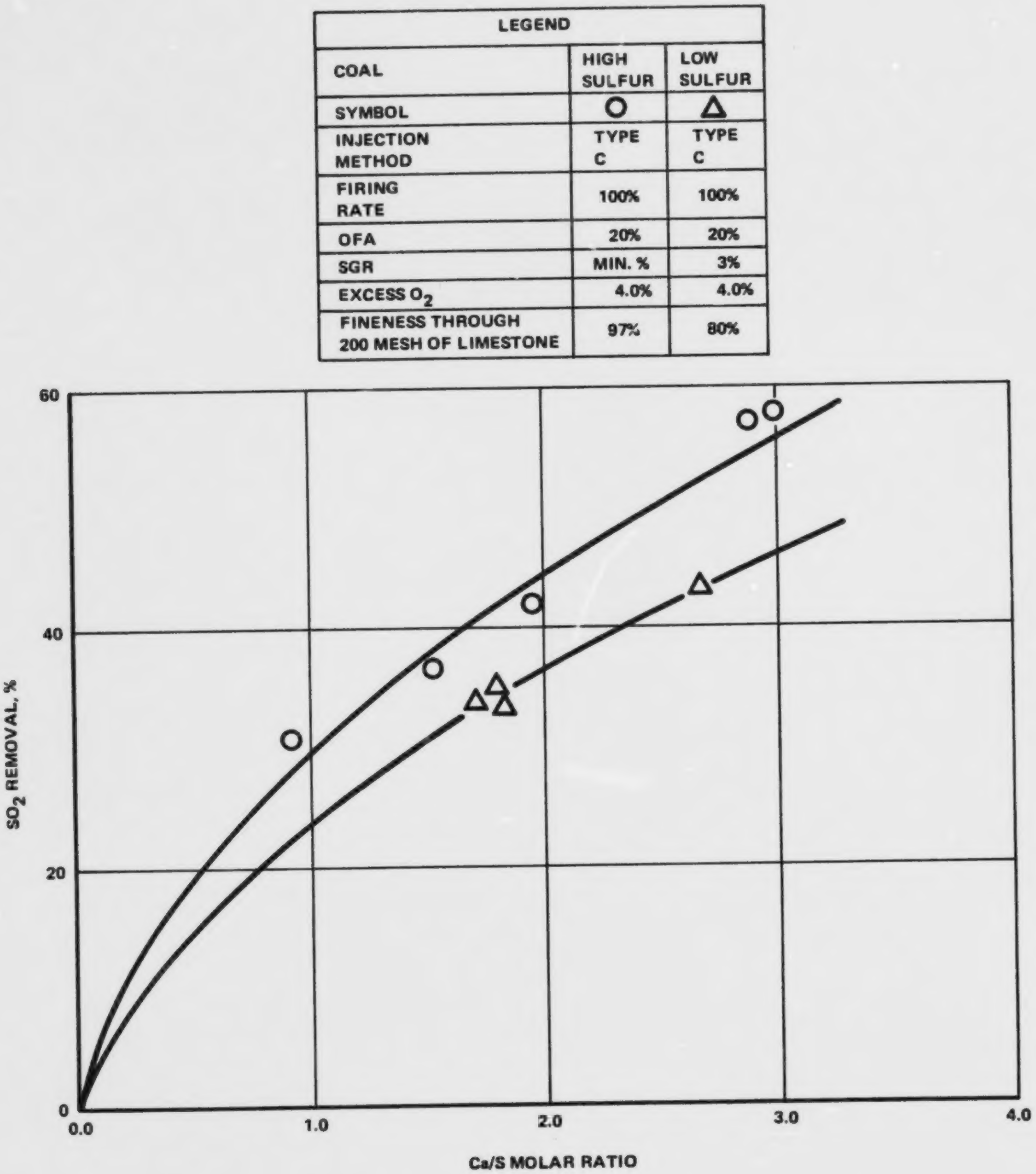


Figure 40. Comparison of SO_2 removal rate with various methods of injection (high sulfur coal) (Ref. 18).



LEGEND		
COAL	HIGH SULFUR	LOW SULFUR
SYMBOL	○	△
INJECTION METHOD	TYPE C	TYPE C
FIRING RATE	100%	100%
OFA	20%	20%
SGR	MIN. %	3%
EXCESS O_2	4.0%	4.0%
FINENESS THROUGH 200 MESH OF LIMESTONE	97%	80%

Figure 41. Effect of Ca/S molar ratio on SO_2 removal (Ref. 18).

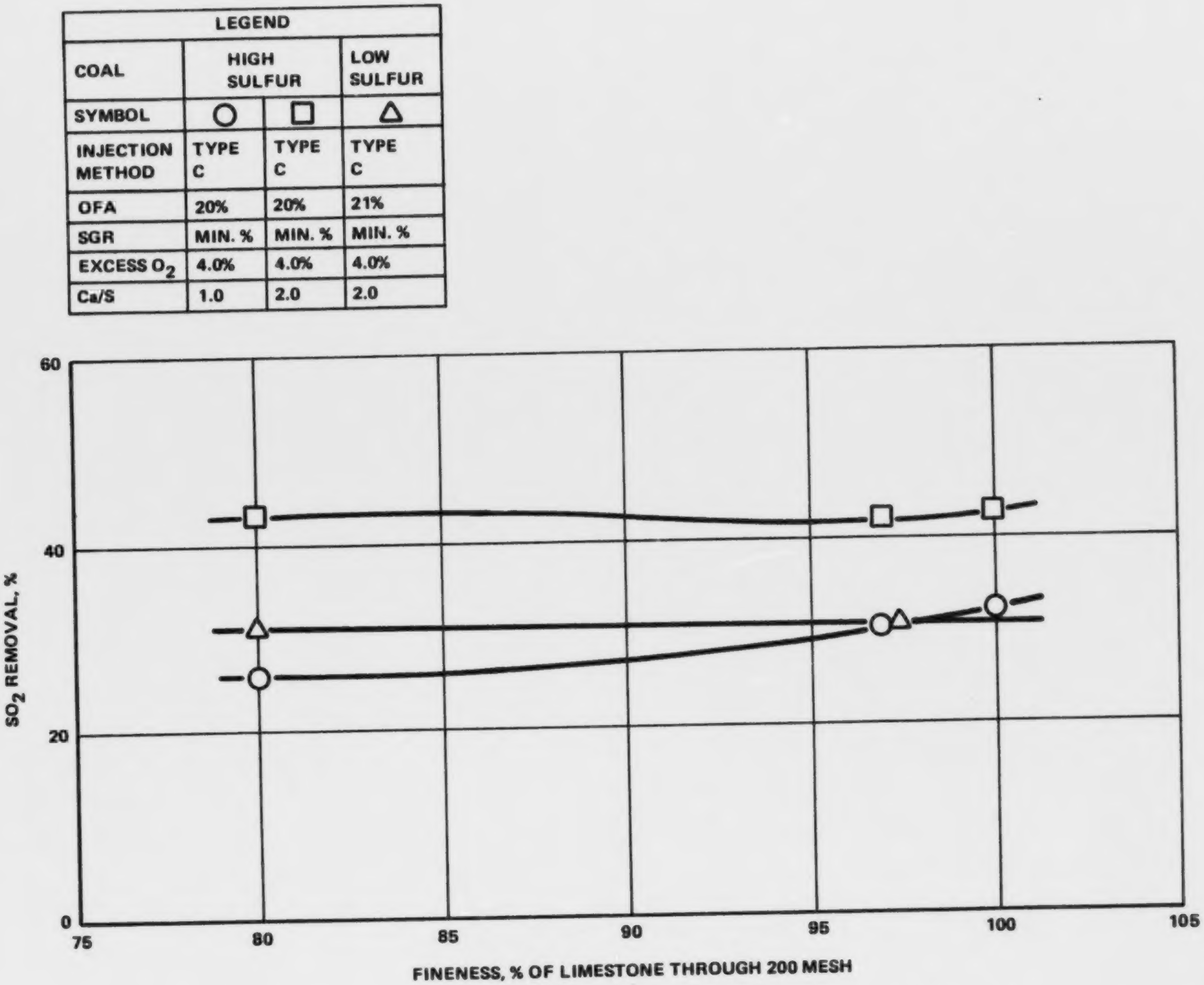


Figure 42. Effect of limestone fineness on SO_2 removal (Ref. 18).

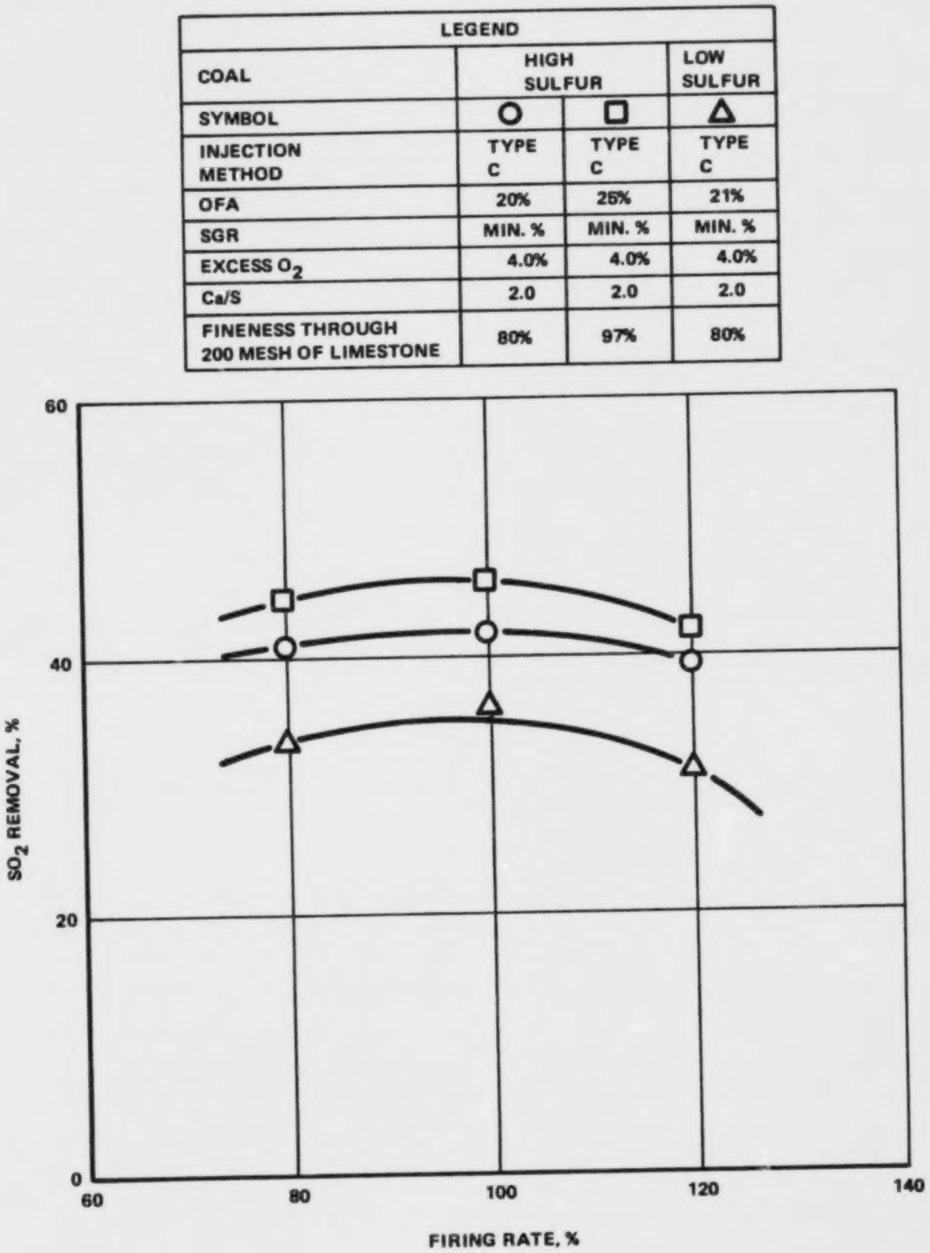


Figure 43. Effect of firing rate on SO_2 removal (Ref. 18).

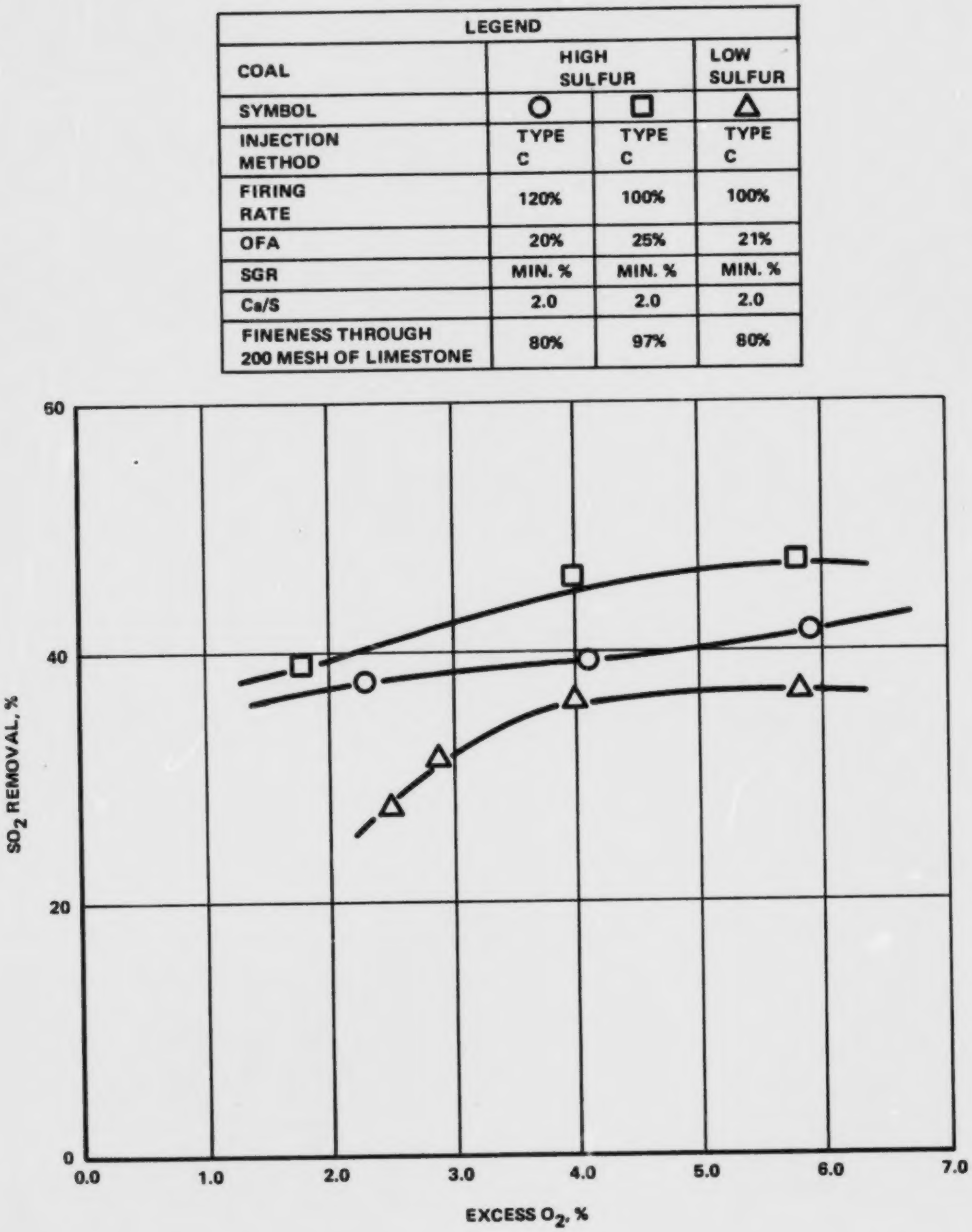
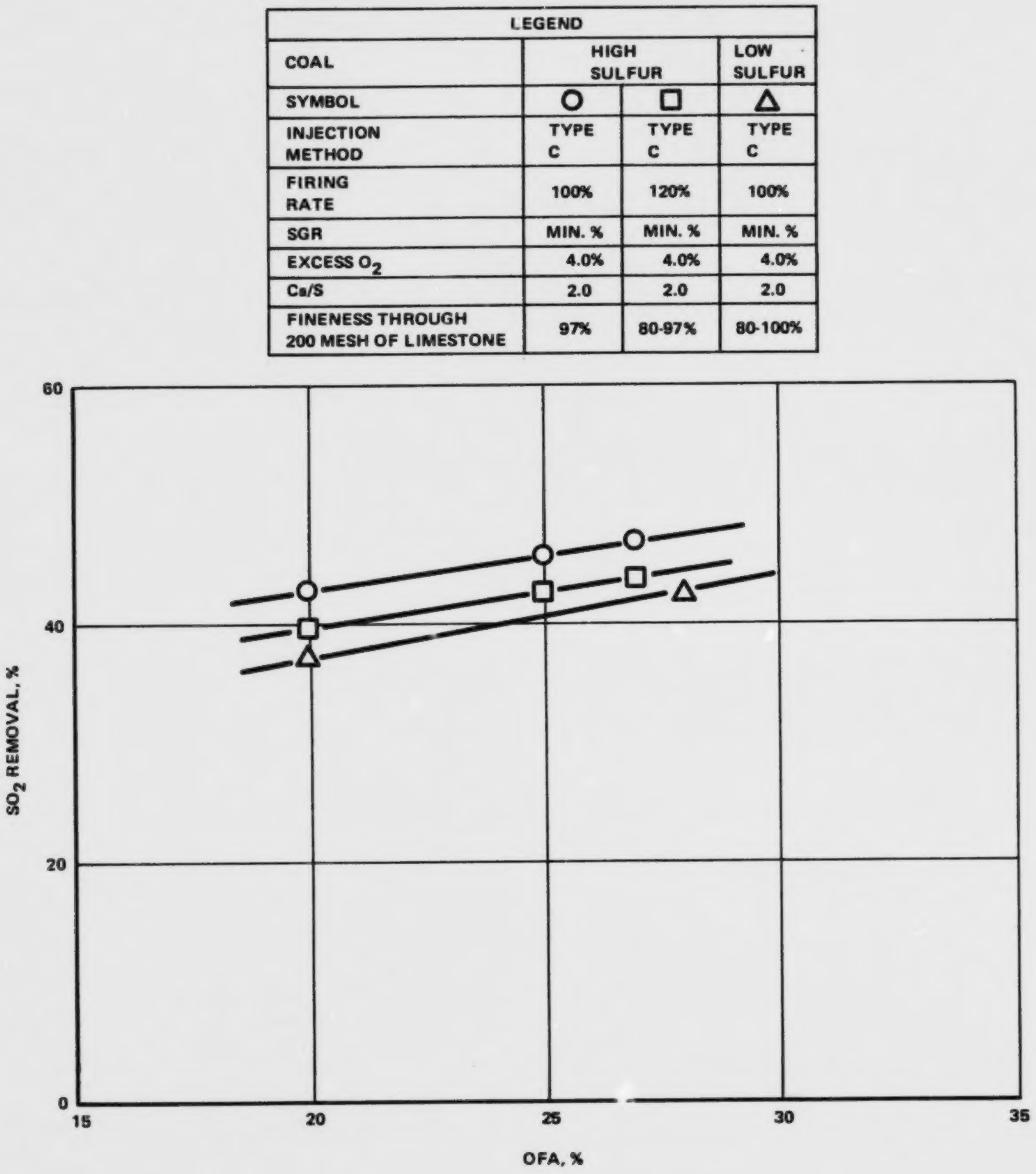


Figure 44. Effect of excess O_2 on SO_2 removal (Ref. 18).



LEGEND			
COAL	HIGH SULFUR		LOW SULFUR
SYMBOL	○	□	△
INJECTION METHOD	TYPE C	TYPE C	TYPE C
FIRING RATE	100%	120%	100%
SGR	MIN. %	MIN. %	MIN. %
EXCESS O_2	4.0%	4.0%	4.0%
Ca/S	2.0	2.0	2.0
FINENESS THROUGH 200 MESH OF LIMESTONE	97%	80-97%	80-100%

Figure 45. Effect of OFA on SO_2 removal (Ref. 18).

improved by increasing the quantity of oxygen in the SO_2 capture reaction zone either by increasing excess oxygen or by increasing the proportion of overfire air.

6.2 Hoot Lake No. 2: 53 MWe Steam Generator

Limestone injection tests were conducted by Radian Corp. at Otter Tail Power Co. Hoot Lake No. 2 station, Ref. (20). This unit is a 53 MWe lignite-fired boiler with tangential burners. The range of calcium to sulfur mole ratios tested varied from 1.6:1 to 5.0:1. Resulting SO_2 removal efficiencies varied from 21% to 47%, as shown in Figure 46. Seven tests were conducted, six of which were in a low-NO_x firing mode. However, it was reported that problems were encountered with grinding of limestone in burner injection tests, resulting in large particles. For the furnace injection tests, the limestone was injected high in the furnace and SO_2 capture may have been limited by poor mixing and short residence time.

6.3 Provence No. 1: 50 MWe Steam Generator

Tests were performed to evaluate the effectiveness of $Ca(OH)_2$ injection for reducing SO_2 emissions at the Provence No. 1 unit in southeastern France, Ref. (25). This unit operates on a high sulfur, high calcium subbituminous coal. The tests were performed to evaluate the effectiveness of SO_2 removal by furnace injection and to optimize the sorbent injection method.

Provence No. 1 is an older 50 MWe unit. The unit has four levels of tilting tangential burners. For the tests the unit was operated at 45 MWe with coal fed only through the lowest three levels and air fed through the upper level to simulate overfire air. The nozzles were operated with zero tilt, and there was 35% air excess at the economizer outlet.

Typical analyses of the Gardanne coal used at Provence are given in Table 7. A significant feature of the Gardanne coal is that the Ca/S mole ratio of the coal itself is always greater than 1.0 and may be as high as 3.0. The calcium in the coal gives removal of 30% to 50% of the sulfur in the coal as shown in Figure 47. This is termed "natural" or "inherent" SO_2 capture.

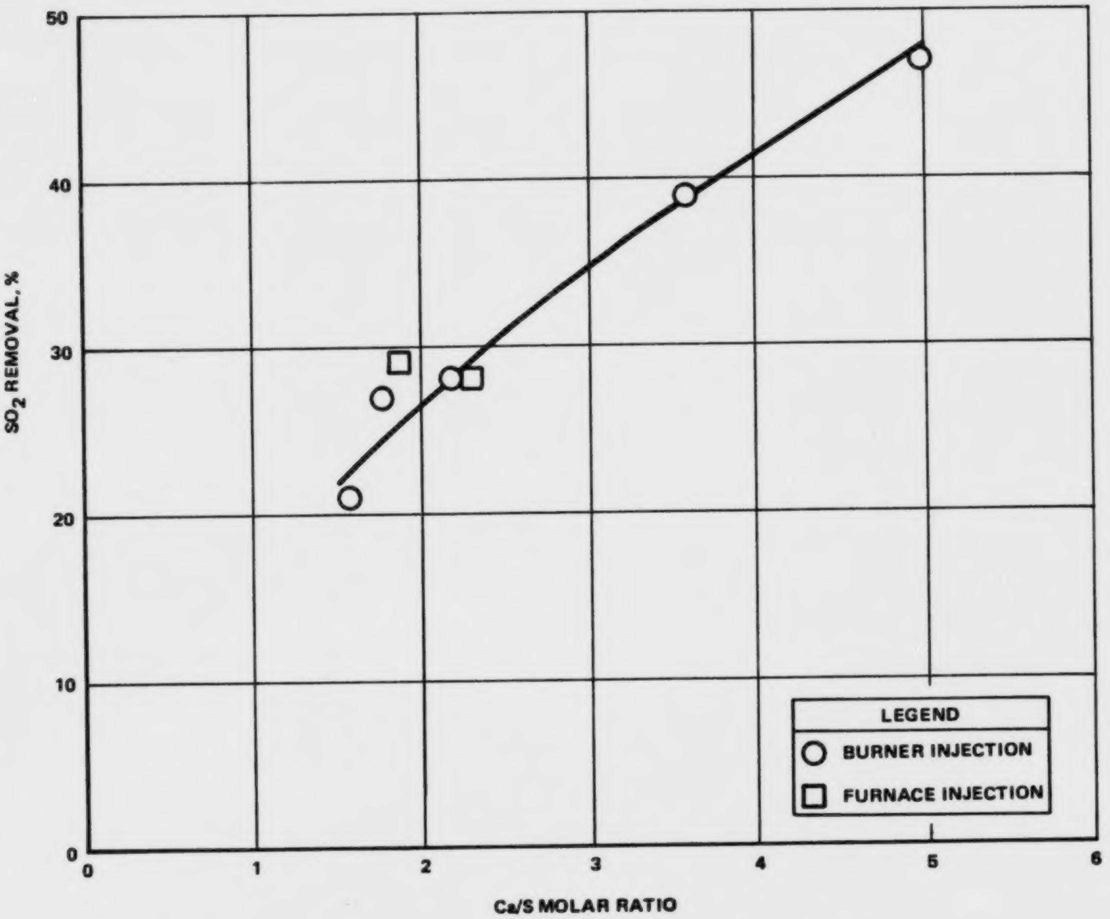


Figure 46. SO_2 removal results for Otter Tail Power Co. limestone injection tests (Ref. 20).

TABLE 7. TYPICAL ANALYSES OF GARDANNE COAL (AS-FIRED BASIS)

	Base Coal	High Ash Coal
Proximate, wt % (including CO_2)		
Moisture	12.0	10.7
Volatile matter	38.2	38.4
Fixed carbon	27.0	22.9
Ash	22.8	28.0
Ultimate, wt % (including CO_2)		
Moisture	12.0	10.7
Ash	22.8	28.0
Carbon	45.7	41.4
Hydrogen	3.1	2.9
Total sulfur	4.1	3.9
Nitrogen	0.9	1.2
Oxygen (diff)	11.4	11.9
Higher heating value, kcal/kg	4360	3790
Higher heating value, Btu/lb	7850	6820
Ash, wt %		
CaO	51.0	59.9
Ca/S molar ratio	1.6	2.5
Sulfur content, kg/10^6 kcal	9.4	10.3
Sulfur content, lb/10^6 Btu	5.2	5.7

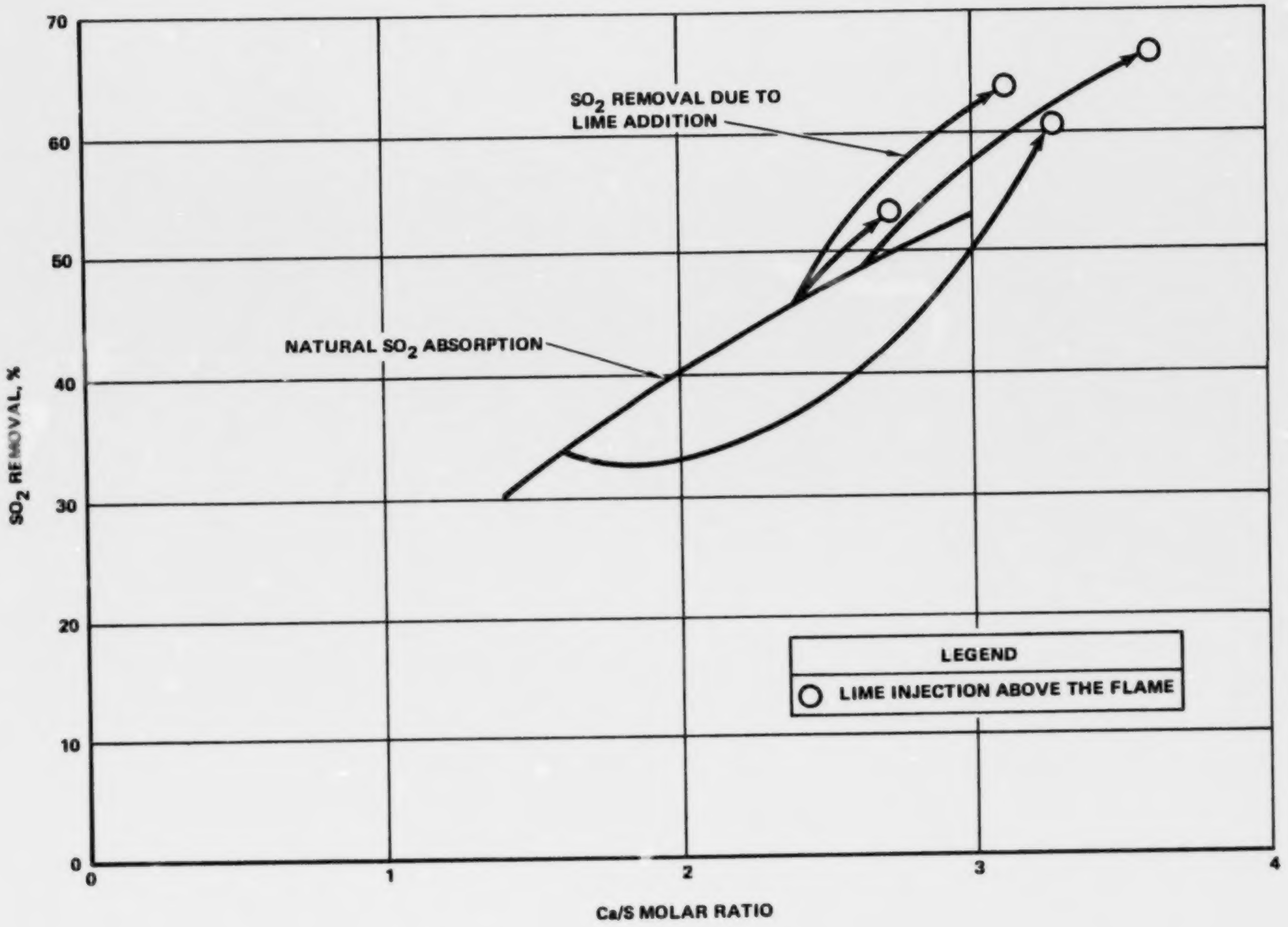


Figure 47. SO_2 removal results for Provence No. 1 hydrated lime injection tests (Ref. 25).

Several furnace operating modes and sorbent injection modes were tested to optimize SO_2 capture. Results were evaluated in terms of the SO_2 capture improvement relative to the natural SO_2 capture found without sorbent injection. Flue gas recirculation without sorbent injection was evaluated to determine whether reducing furnace temperature would be effective. No improvement in SO_2 relative to natural SO_2 capture was measured. Similarly, reducing furnace temperature by injection of water into the furnace gave no improvement in SO_2 capture.

Tests were performed with $Ca(OH)_2$ fed directly into the three operating coal pulverizers and into the furnace through injectors mounted close to and parallel to the 12 operating coal nozzles. Negligible improvement in SO_2 capture was measured relative to the natural SO_2 capture. On the other hand, injection of $Ca(OH)_2$ above the flame through the unused upper coal elevation gave a significant improvement in SO_2 removal compared to the natural SO_2 removal. These results are shown in Figures 47 and 48. The tests were run both without upper level secondary air and with secondary air surrounding the additive feed. As shown in Figure 48, improvement in SO_2 capture was obtained with the secondary air on. This improvement was attributed to improved penetration and mixing of the additive stream which existed when the upper level secondary air was in operation.

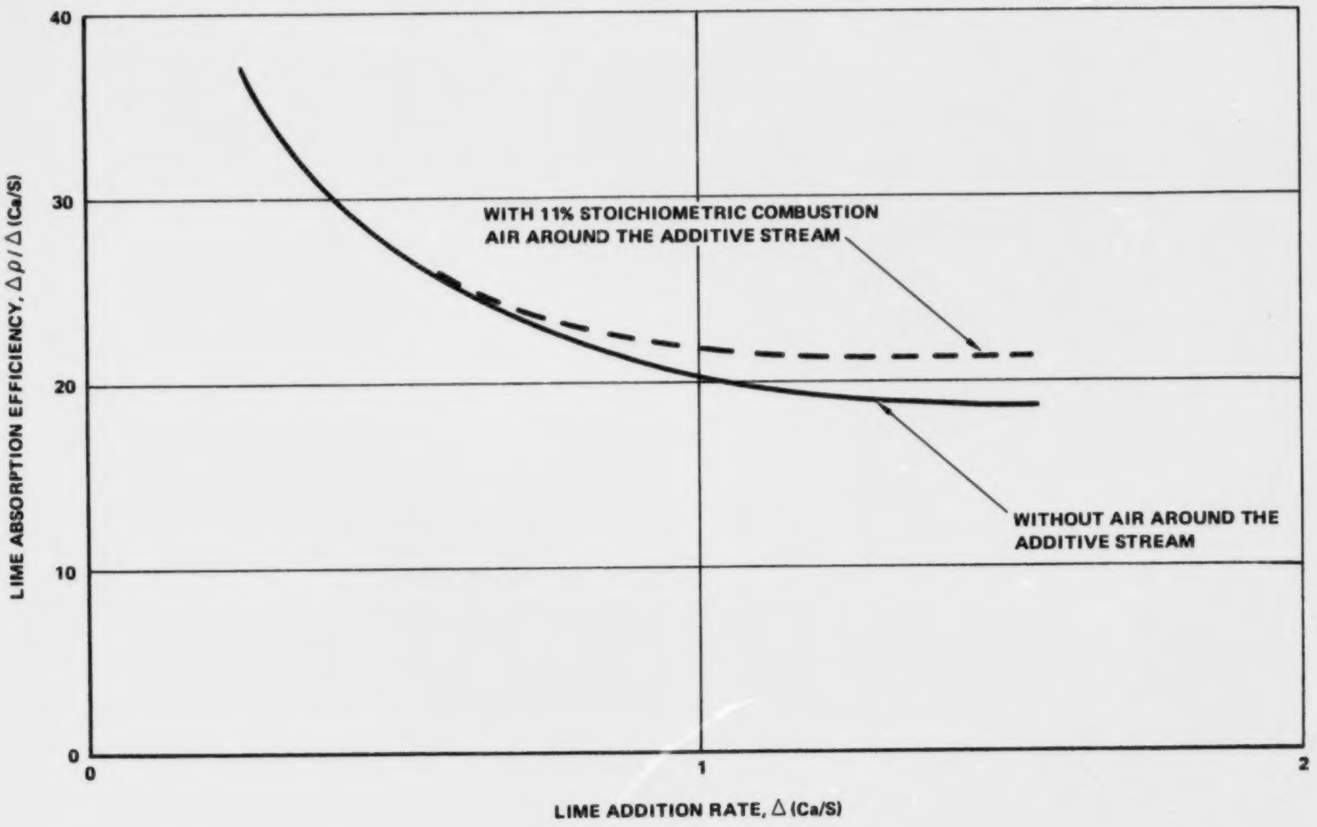


Figure 48. Improvement in lime absorption efficiency with addition of secondary air (Provence No.1) (Ref. 25).

SECTION 7

EFFECTS OF SORBENT INJECTION ON FLUE GAS CLEANING

Addition of limestone to a utility boiler may also affect the operation of downstream flue gas cleaning equipment. The particular effects will tend to be equipment-specific, but relate to the fact that more particulate is produced, and that the chemical composition of the total particulate matter is changed. Areas and equipment to be discussed include bottom ash and fly ash removal and disposal, collection efficiency of baghouses and electrostatic precipitators, and effects on, and potential integration with, tail-end sulfur dioxide removal systems.

7.1 Ash Collection and Disposal

Large-scale testing at TVA revealed potential problems with limestone injection in the areas of bottom ash removal and electrostatic precipitator performance. With respect to bottom ash removal, the flyash/calcined limestone mixture caused hydraulic conveyor plugging due to its more cementitious properties compared to flyash alone. Smooth surface coating or increased flushing frequency did not correct deposits at the wet/dry interface.

Electrostatic precipitator performance was a function of limestone particle size, flue gas temperature, coal sulfur, and limestone injection rate, but, in general, was detrimentally affected by additive injection. In contrast, performance of the mechanical collector was relatively insensitive to all conditions of injection or non-injection and ranged in efficiency between 50% and 60%.

In the B & W pilot-scale tests, electrical resistivity of the flyash produced from different coals and additives was determined in situ and also in a laboratory test chamber. All limestone and dolomite samples caused a high resistivity situation. It was concluded that performance of an electrostatic precipitator would decrease with sorbent injection.

At Detroit Edison, dust loading in the stack gas increased an average of about 65% during tests conducted by Combustion Engineering with dolomite injection. Mechanical collectors upstream of the precipitator operated with no loss of efficiency, and particle size distribution was determined to be about the same as normal flyash. However, performance of the electrostatic precipitator was significantly affected by the change in particulate composition and loading. The electrical characteristics changed almost immediately from loaded to sparking conditions and collection efficiency dropped from about 80% (normal) to 55% at an operating temperature of 270°F. Overall collection efficiency, (mechanical plus electrostatic), dropped from 96% to 90%.

For retrofit applications, some improvement in ESP performance can be realized by SO_3 conditioning. Tests have shown that LIMB flyash resistivity can be significantly reduced by injecting small amounts of SO_3 ($\sim$ 5 ppm), increasing the flue gas moisture content, and/or decreasing the outlet temperature to less than 250°F. Additional research is needed in this area before the extent of ESP modification can be accurately defined.

With sorbent injection into the furnace, all ash hoppers will likely experience a higher loading. In general, ash hoppers are oversized to accommodate a range of ash loadings from various coals. It is reasonable to assume that no modifications will be required to compensate for additional solids collection. Approximately 20% to 30% of total solids are typically collected in the hoppers. For this study, it was assumed that all of the sorbent-related solids would pass through the boiler to the back end cleanup equipment. If a significant increase in solids collected in the hoppers is experienced, the operational frequency of the ash removal systems may have to be increased.

When the ash from a limestone-injected boiler is collected dry with an ESP or bag filter, it is desirable to avoid contact with water until the material is conveyed to its ultimate disposal. Dry/wet interfaces in the ash removal system may be a source of plugging or deposit buildup which would reduce the capacity of the ash conveying system. A completely dry ash removal and conveying system would be optimal.

Past experience with wet scrubbers shows that limestone ash, removed in a sluice system, behaves much like coal ash. It may be slurried and conveyed with no unusual problems of agglomeration. Any potential settling or cementation problems could be prevented by flushing the slurry lines during unit outages.

7.2 Integration of Sorbent Injection and Tail End Scrubbing

A major concern in the design of a system using both furnace injection and tail end scrubbing is the impact of the furnace calcined additive on scrubber chemistry. In previous system designs, the wet scrubber was used as both the particulate collection device and SO_2 absorber, directly receiving all of the sorbent injection into the furnace. These highly reactive alkaline earth oxides can create chemical scaling problems in certain absorber designs under particular operating conditions. Specifically, an excess of sorbent material introduced into a scrubbing system can cause $CaSO_3$ scaling or even $CaCO_3$ scaling if liquor pH's get very high. In a packed-bed absorber, scaling can necessitate shutting down the system for cleaning. For an open absorber design, such as a spray tower, scaling could still occur, but would be less likely to require manual cleaning, since the deposits would redissolve once normal operation is resumed.

However, in today's common wet scrubbing designs, an electrostatic precipitator or baghouse is placed ahead of the SO_2 absorber for particulate collection. In this case, the calcinated material would be collected before the absorber and could be reslurried and fed into the scrubber at a controlled rate. Absorber design and careful regulation of sorbent feed would eliminate the scaling problems encountered previously in furnace injection-wet scrubbing systems.

With regard to furnace injection combined with other tail end scrubbing technologies, the concept of "dry scrubbing" (i.e., SO_2 absorption in a spray dryer) is rapidly becoming a preferred method for certain situations and also offers unique advantages in a combined system approach. Dry scrubbing alone, compared to wet tail end limestone scrubbing, is generally acknowledged to have lower capital and operating costs for low-to-medium sulfur coal applications. For high sulfur coals, the increased sorbent requirements of dry scrubbing, coupled with much higher sorbent costs (lime vs. limestone), result in wet scrubbing being economically more favorable. However, the CaO produced by furnace injection of limestone is potentially suitable as an absorbent for dry scrubbing. In a combined system, the furnace calcined material could be recycled from the dry scrubber bottom or fabric filter for wet injection into the dry scrubber, eliminating the need to purchase commercial lime. Operating costs would therefore be reduced significantly.

In addition to economics, both furnace injection and dry scrubbing used alone have difficulty meeting NSPS removal requirements with high sulfur coals. Used in a combined mode, their application may be extended to these coals, also. For low or medium sulfur coals, partial treatment of the flue gas by dry scrubbing could be employed where in-furnace removal alone was not adequate to meet emission requirements.

High efficiency SO_2 absorption in a spray dryer has been shown to require more reactive alkalies, such as $Ca(OH)_2$ or sodium alkalies, than wet scrubbing. An important question regarding merging of the two technologies is reactivity of the furnace-calcined limestone in a spray dryer SO_2 absorber. In the late 1960's, various trials of limestone furnace injection, followed by wet scrubbing, revealed poor CaO reactivity in the scrubbers, presumably due to deadburning. However, limestone calcined in a boiler with lower temperatures due to low-NO_x firing may produce more reactive lime.

In recent testing by Combustion Engineering for EPRI, furnace-calcined limestone was evaluated as an additive for SO_2 absorption in a dry scrubbing system. Figure 49 is a schematic of the test system, which extracted a slip stream of about 100 acfm at 350°F from a 3 to 4 x 10^6 Btu/hr furnace firing coal under low-NO_x conditions with limestone injection. Tests were conducted using dry material collected from the baghouse as additive to the atomizer of the spray dryer. The additive was slaked and slurried before being fed. Commercial lime was also used as the spray dryer additive for a portion of the tests to provide a basis for comparison.

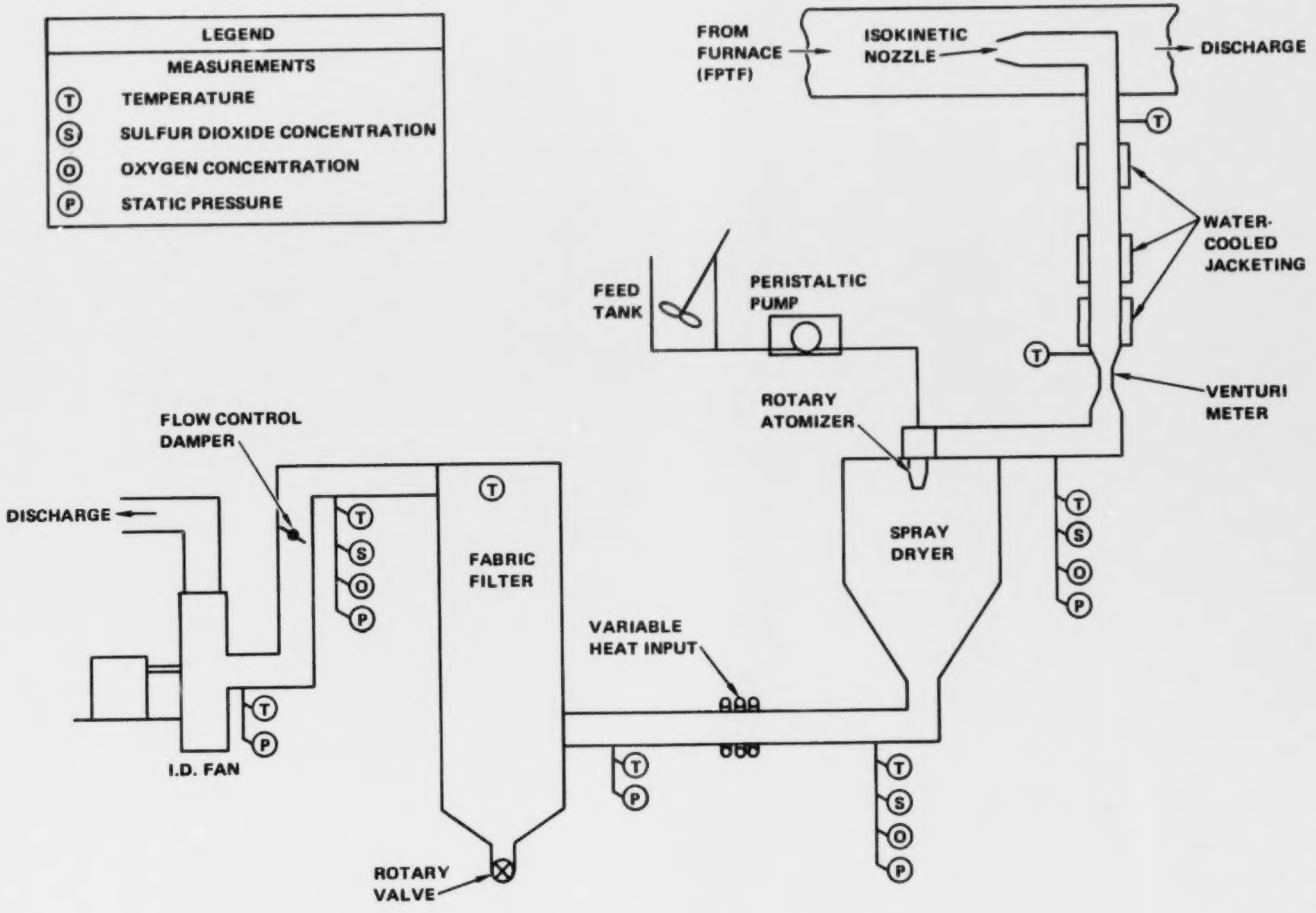

LEGEND
MEASUREMENTS
T TEMPERATURE
S SULFUR DIOXIDE CONCENTRATION
O OXYGEN CONCENTRATION
P STATIC PRESSURE
FROM FURNACE (FPTF)
ISOKINETIC NOZZLE
DISCHARGE
T
WATER-COOLED JACKETING
VENTURI METER
FEED TANK
PERISTALTIC PUMP
ROTARY ATOMIZER
SPRAY DRYER
T S O P
VARIABLE HEAT INPUT
T P
T S O P
FABRIC FILTER
T
ROTARY VALVE
FLOW CONTROL DAMPER
DISCHARGE
T S O P
T P
I.D. FAN

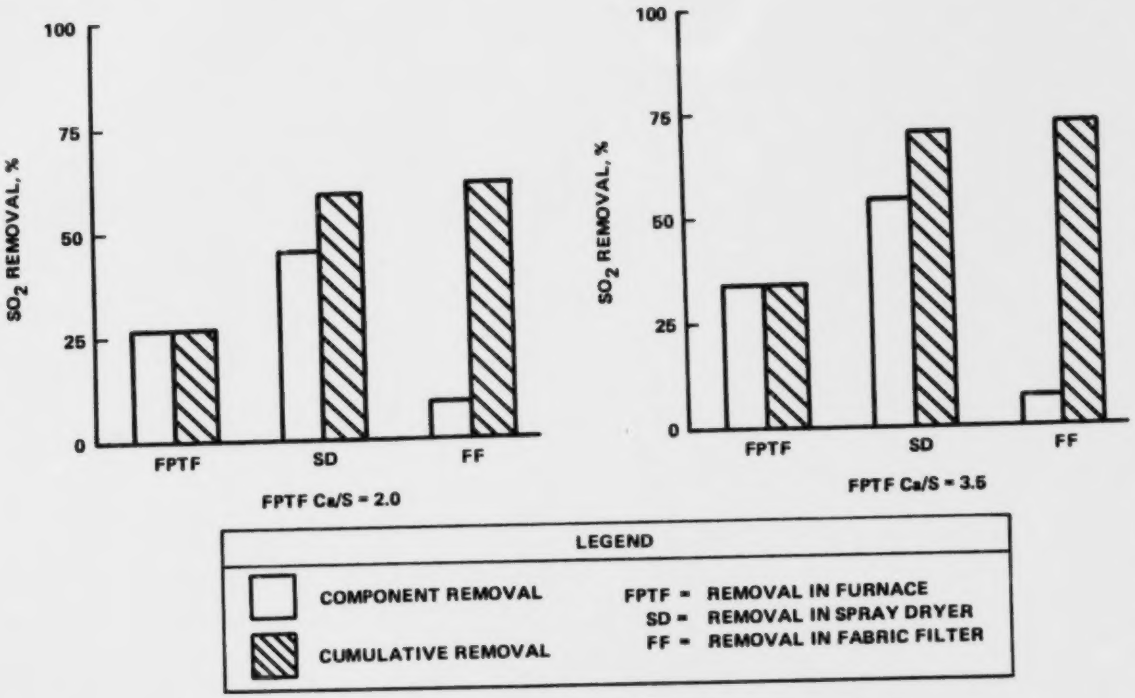
Figure 49. Schematic of the dry scrubbing test facility.

On the average, furnace-calcined lime was determined to produce SO_2 removal efficiencies of about one-half (0.5) those produced by commercial lime at equivalent stoichiometry and similar operating conditions. These results were in the absence of dry, furnace-calcined limestone entering the spray dryer with the flue gas. With furnace-calcined limestone in the flue gas entering the spray dryer, its contribution to SO_2 removal increased overall removal efficiency to 0.7 to 0.8 of that obtained with commercial lime.

In Figure 50, average SO_2 removal efficiencies are shown for the entire system of furnace, spray dryer, and fabric filter for limestone stoichiometries of 2:1 and 3.5:1 fed to the furnace. The stoichiometry of recycled additive atomized to the spray dryer was 1:1, and was limited to this value due to the slurry handling capacity of the small atomizer. Larger systems could operate at higher stoichiometries and, therefore, higher removal efficiencies.

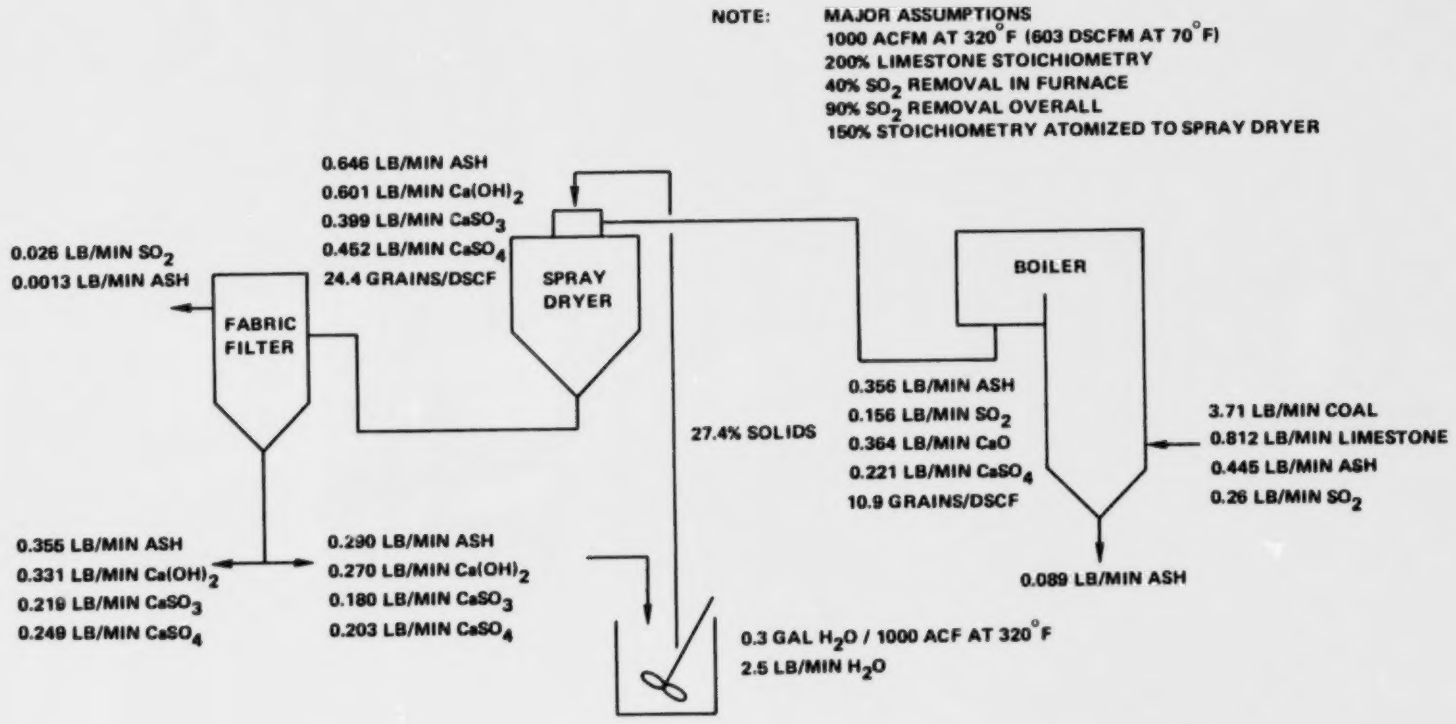
One of the differences between a dry scrubbing system using commercial lime and one using furnace-calcined material is the amount of flyash which must be handled in the spray dryer circuit. To illustrate flows and compositions for gas, solid, and liquid streams in a dry scrubbing/furnace injection system, a material balance diagram from the EPRI final report is shown in Figure 51. Assumptions used to define the material balance are as follows:

- coal composition: 3.5% sulfur; 12% ash; 12,000 Btu/lb
- flow basis of 1000 acfm @ 320°F
- 200% limestone stoichiometry fed to boiler (Ca/S of 2:1)
- 40% SO_2 removal in furnace
- 150% reagent ratio fed to spray dryer atomizer (not including alkali in flue gas)
- 83% SO_2 removal in spray dryer + fabric filter
- 90% SO_2 removal overall
- 20% ash drop out in boiler



NOTES:
1. COMPONENT REMOVALS ARE CALCULATED BASED ON THE INLET SO_2 CONCENTRATION TO THAT COMPONENT.
2. CUMULATIVE REMOVALS ARE BASED ON THE SO_2 CONCENTRATION WHICH WOULD EXIT THE FPTF IF NO LIMESTONE WAS INJECTED INTO THE FURNACE.
3. BOTH GRAPHS ARE FOR RECYCLED CALCINED MATERIAL FED TO THE SPRAY DRYER AT A REAGENT RATIO OF APPROXIMATELY 100%.

Figure 50. Distribution of SO_2 removal efficiencies for the FPTF and DSTF.


NOTE: MAJOR ASSUMPTIONS
1000 ACFM AT 320° F (603 DSCFM AT 70° F)
200% LIMESTONE STOICHIOMETRY
40% SO_2 REMOVAL IN FURNACE
90% SO_2 REMOVAL OVERALL
150% STOICHIOMETRY ATOMIZED TO SPRAY DRYER

FABRIC FILTER
SPRAY DRYER
BOILER
0.026 LB/MIN SO_2
0.0013 LB/MIN ASH
0.646 LB/MIN ASH
0.601 LB/MIN $Ca(OH)_2$
0.399 LB/MIN $CaSO_3$
0.452 LB/MIN $CaSO_4$
24.4 GRAINS/DSCF
0.356 LB/MIN ASH
0.156 LB/MIN SO_2
0.364 LB/MIN CaO
0.221 LB/MIN $CaSO_4$
10.9 GRAINS/DSCF
3.71 LB/MIN COAL
0.812 LB/MIN LIMESTONE
0.445 LB/MIN ASH
0.26 LB/MIN SO_2
0.089 LB/MIN ASH
27.4% SOLIDS
0.355 LB/MIN ASH
0.331 LB/MIN $Ca(OH)_2$
0.219 LB/MIN $CaSO_3$
0.249 LB/MIN $CaSO_4$
0.290 LB/MIN ASH
0.270 LB/MIN $Ca(OH)_2$
0.180 LB/MIN $CaSO_3$
0.203 LB/MIN $CaSO_4$
0.3 GAL H_2O / 1000 ACF AT 320° F
2.5 LB/MIN H_2O

Figure 51. Material balance for dry scrubbing/furnace limestone injection system.

- flyash removal in fabric filter to attain outlet of 0.3 lbs ash/10^6 Btu
- 10% oxidation of sulfite to sulfate in spray dryer reaction products
- $CaSO_3$ and $CaSO_4$ are unhydrated

The SO_2 removal performance of the dry scrubbing/furnace injection system was specified to meet NSPS limits, and demonstrate whether stream flows, concentrations, dust loadings, etc., are technically feasible at this level of performance. Although the SO_2 removals shown are optimistic compared to the pilot results, commercial-scale performance of the spray dryer/fabric filter is expected to be better for the reason mentioned.

Ash loadings to the fabric filter downstream of the spray dryer could be considerable, although the loadings will vary depending on the spray dryer design (with or without solids discharge). In Figure 51, all solids are shown transported to the fabric filter, producing a loading of 24.4 grains/DSCF. Dry solids dropout of approximately 50% has been observed in spray dryers with a solids discharge, and would decrease the loading to the fabric filter accordingly. For the dry scrubbing/furnace injection system, separate solids discharge from the spray dryer would be advantageous.

SECTION 8

HISTORICAL AND PROJECTED DESIGN TRENDS

An investigation was conducted to analyze the evolution of various design characteristics of Combustion Engineering, Inc.'s tangentially fired boilers. The results are intended to serve as guidelines for establishing criteria for site selection and for identifying those boiler design parameters which are influential in designing for dry sorbent injection.

To provide an understanding of boiler design trends, a survey of key design parameters for C-E coal-fired boilers was undertaken. The survey covered a total of 273 units sold from 1960 to 1982. Of this total, 181 units, built for domestic utilities, were designed for coal firing only. The remaining units were either built overseas, designed for multifuel capability, delayed, or cancelled. This breakdown is shown in Table 8. The 179 coal-only units sold between 1960 and 1980 were used as the basis for study during this analysis.

The following design parameters were considered:

1. Plant size (MWe)
2. Coal rank
3. Unit plan area and height
4. Unit net heat input per plan area (Btu/hr-ft^2) and volumetric heat release rate (Btu/hr-ft^3).
5. Number of burners per unit
6. Heat input per burner (Btu/hr)
7. Number of sootblowers (wall and retractable)
8. Air quality control system

These parameters are presented graphically as a function of contract year and/or unit size (MWe). The effect of coal range on some parameters is illustrated. Descriptions of each parameter and graph are presented in the following sections.

TABLE 8. C-E COAL FIRED UTILITY UNITS SOLD DURING THE PERIOD 1960-1982

	1960-1980	1981-1982	Total
Coal only	179	2	181
Multi-fuel	44	-	44
Overseas	28	8	36
Delayed	4	2	6
Canceled	6	-	6
Total	261	12	273

8.1 Plant Size (MWe)

Plant size (MWe) is plotted against contract year in Figure 52. The data can be roughly broken down into three time periods. In the 1960-1965 era, the majority of the units built were in the range of 150-600 MWe, with three units at 900 MWe. The average size was approximately 400 MWe. From 1966 to 1975, the units built were in the range of 250-850 MWe, with two units at 900 MWe. The average size was approximately 600 MWe. In the 1976 to 1980 time span, the majority of units built were in the range of 350-750 MWe, with an average size of approximately 550 MWe. These data show that the recent trend over the last several years has been toward somewhat smaller units. Prior to 1976, 20 units of 800 MWe or greater were built. Since than, no units above 750 MWe have been built.

Figure 53 shows these units broken down into four groups expressed as a percentage of the total sample. The four groups (classes) and their respective percentages are: 100-300 MWe - 11%; 300-500 MWe - 29%; 500-700 MWe - 36%, 700-900 MWe - 24$. A total generating capacity of 96,322 MWe is represented in this grouping.

8.2 Coal Rank

There are five basic types or ranks of coal which have been used during the past 25 years and which will continue to be used during at least the rest of the century. Typical analyses for these coals are shown in Table 9. As can be seen, there is a wide variation in heating value as well as moisture, ash, and sulfur content. The ash analyses also show substantial differences in the amounts of iron and sodium oxides in the ash. Differences in the amounts of these constituents significantly influence the design and operation of a boiler.

Figure 54 shows the selected sample of boilers segregated by primary coal fired and contract year. It can be seen that during the period from 1960 to 1970, most units were designed to burn Eastern or Midwestern bituminous coal. These coals represented roughly 45% and 30% of the total (C-E) market,

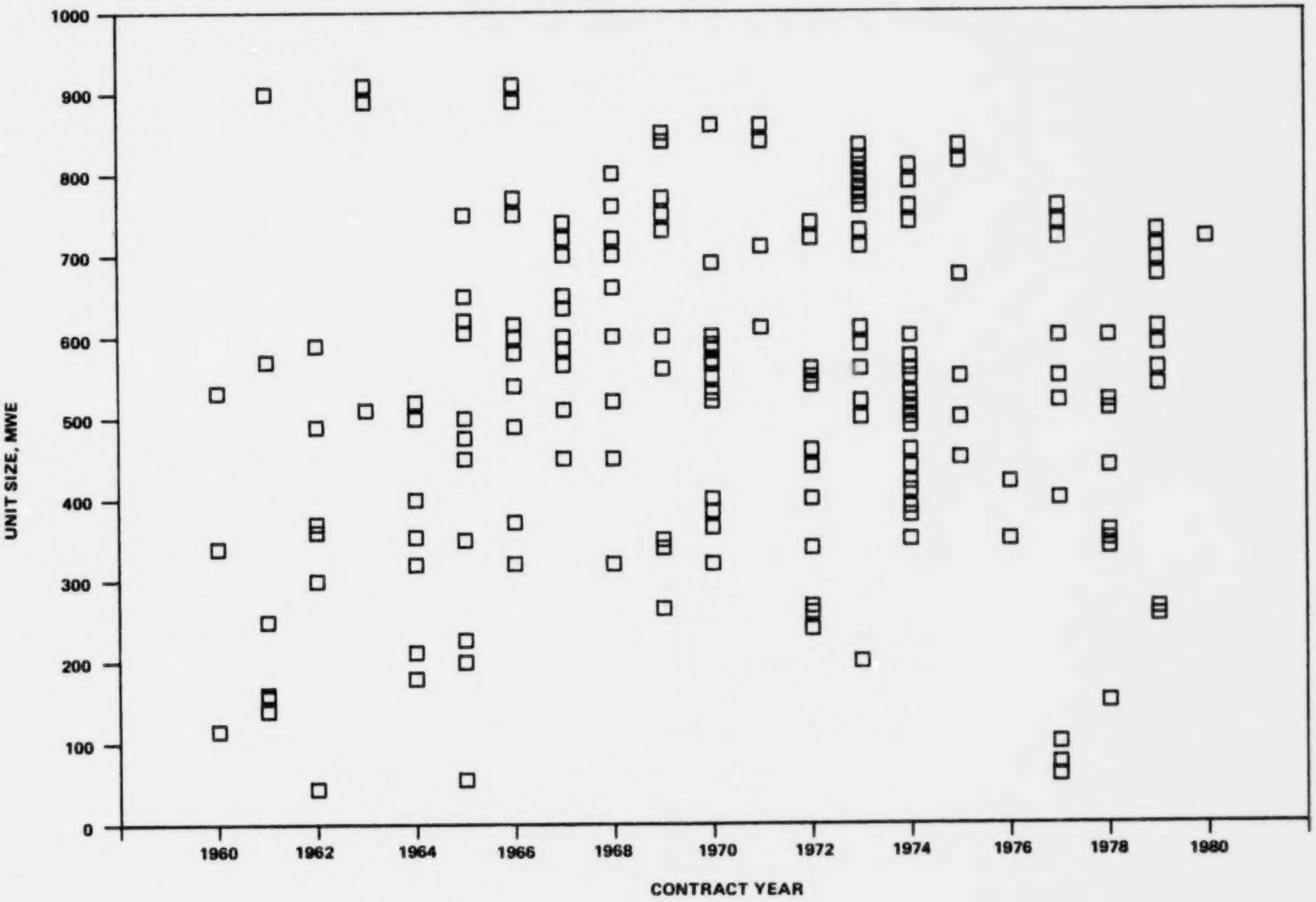


Figure 52. Domestic coal-fired utility boilers sold by C-E between 1960 and 1980.

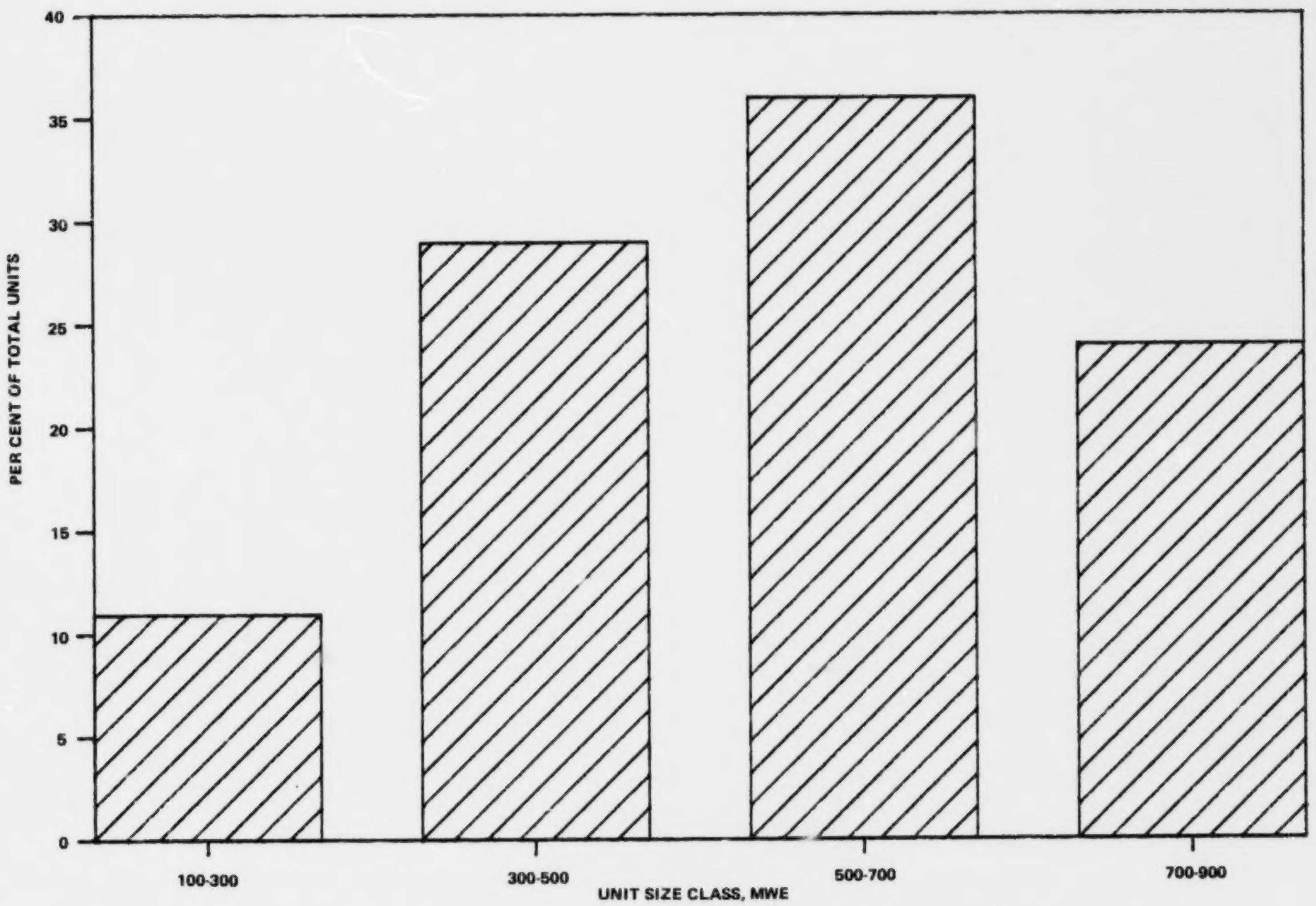


Figure 53. Distribution of C-E domestic utility coal-fired boilers by unit size (1960-1980).

TABLE 9. TYPICAL COAL ANALYSES

	Eastern Bituminous	Midwest Bituminous	Sub-bituminous C	Texas Lignite	Northern Plains Lignite
Total moisture, %	5.0	15.4	30.0	31.0	39.6
Ash, %	10.3	15.0	5.8	10.4	6.3
Volatile matter, %	31.6	33.1	32.6	31.7	27.5
Fixed carbon, %	53.1	36.5	31.6	26.9	26.6
Total %	100.0	100.0	100.0	100.0	100.0
Btu/lb, as-fired	13,240	10,500	8,125	7,590	6,523
Btu/lb, moisture & ash free	15,640	15,100	12,650	12,940	12,050
Fusion temperature (reducing), °F					
Initial deformation (ID)	2,170	1,990	2,200	2,075	2,027
Softening (ST)	2,250	2,120	2,250	2,200	2,089
Fluid (FT)	2,440	2,290	2,290	2,310	2,203
Ash analysis, %					
SiO_2	40.0	46.4	29.5	46.1	23.11
Al_2O_3	24.0	16.2	16.0	15.2	11.29
Fe_2O_3	16.8	20.0	4.1	3.7	8.48
CaO	5.8	7.1	26.5	16.6	23.75
MgO	2.0	0.8	4.2	3.2	5.87
Na_2O	0.8	0.7	1.4	0.4	7.38
K_2O	2.4	1.5	0.5	0.6	0.70
TiO_2	1.3	1.0	1.3	1.2	0.45
P_2O_5	0.1	0.1	1.1	0.1	-
SO_3	5.3	6.0	14.8	12.7	17.69
Not accounted for	1.5	0.2	0.6	0.2	1.28

TABLE 9. TYPICAL COAL ANALYSES - CONTINUED

	Eastern Bituminous	Midwest Bituminous	Sub-bituminous C	Texas Lignite	Northern Plains Lignite
Sulfur,%	1.8	3.2	0.34	0.6	0.75
Lb moisture/10^6 Btu	3.8	14.7	36.9	40.8	60.7
Lb ash/10^6 Btu	7.8	14.3	7.1	13.7	9.7
Lb sulfur/10^6 Btu	1.36	3.05	0.42	0.79	1.15
Base/acid ratio	0.426	0.473	0.784	0.392	1.33
Hardgrove grindability	55	56	43	48	25

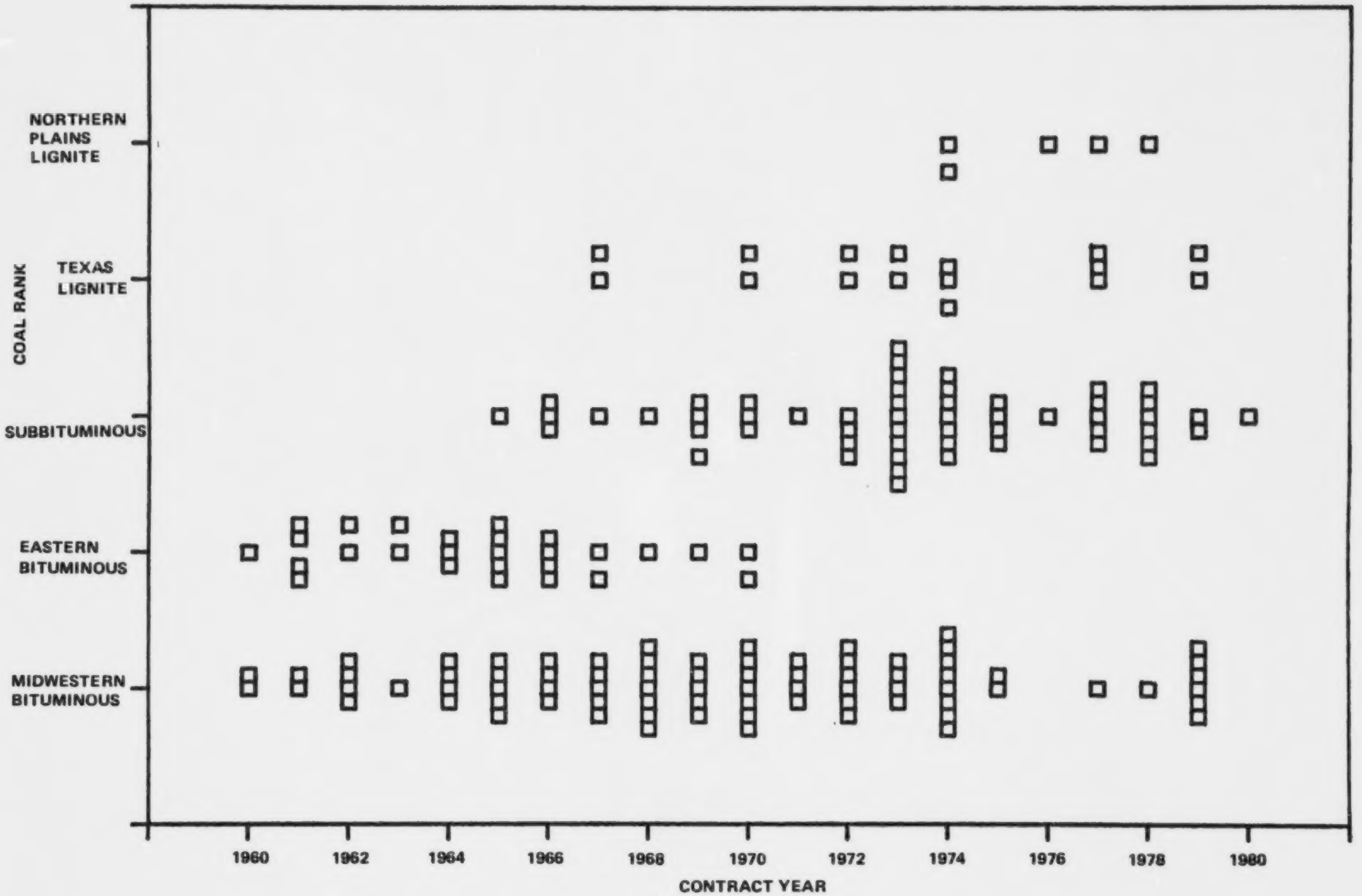


Figure 54. Distribution of C-E domestic utility coal-fired boilers by coal rank and contract year (1960-1980).

respectively. After 1970, the primary coal of choice was sub-bituminous C (50%), followed by Midwestern bituminous (30%), and lignite (20%). The factors most directly responsible for effecting this change in coal fired were the advent of more stringent federal emission regulations in 1970 and industrial expansion of the western and southern United States. The low sulfur content of the sub-bituminous and lignite coals make them more attractive from a capital outlay standpoint, since this greatly reduces the amount which must be spent on expensive cleanup equipment. Table 10 gives the distribution of units by coal range and total generating capacity.

8.3 Unit Plan Area and Height

The most important step in designing a coal-fired unit is to properly size the furnace. As shown in Table 9, the various coals differ significantly in their ash characteristics. Among the properties considered most important are ash chemical composition and ash content (lb ash/million Btu). These characteristics and others translate into the furnace sizes shown in Figure 55. This generalized figure illustrates how furnace size, expressed in terms of plan area, volume, and fuel burning zone height (shown by "h" in Figure 55) required to generate an equivalent amount of steam, increases as lower rank fuels are fired.

The furnace plan area is determined by thorough evaluation of all fuels expected to be fired in the unit and arriving at a desired value of net heat input per plan area (NHI/PA, Btu/hr-ft^2).

The fuel burning zone height depends on the unit width and depth, in addition to the fuel-ash properties. This distance varies from 45 ft to 60 ft in current designs.

Furnace dimensions must be adequate to provide the necessary furnace retention time to properly burn the fuel as well as to cool the gaseous combustion products. This will insure that the final furnace outlet gas temperature at the entrance to the closely spaced convective surfaces is well below the ash softening temperature of the lowest quality coal burned.

TABLE 10. DISTRIBUTION OF COAL-FIRED UTILITY UNITS BY COAL RANK AND TOTAL GENERATING CAPACITY (1960-1980)

Coal Rank	Number of Units	Total MW	% of Total
Eastern bituminous	27	15,071	15.6
Midwestern bituminous	76	37,604	39.0
Subbituminous	55	31,252	32.5
Norther Plains lignite	5	1,920	2.0
Texas lignite	16	10,475	10.9
Total	179	96,322	100.0

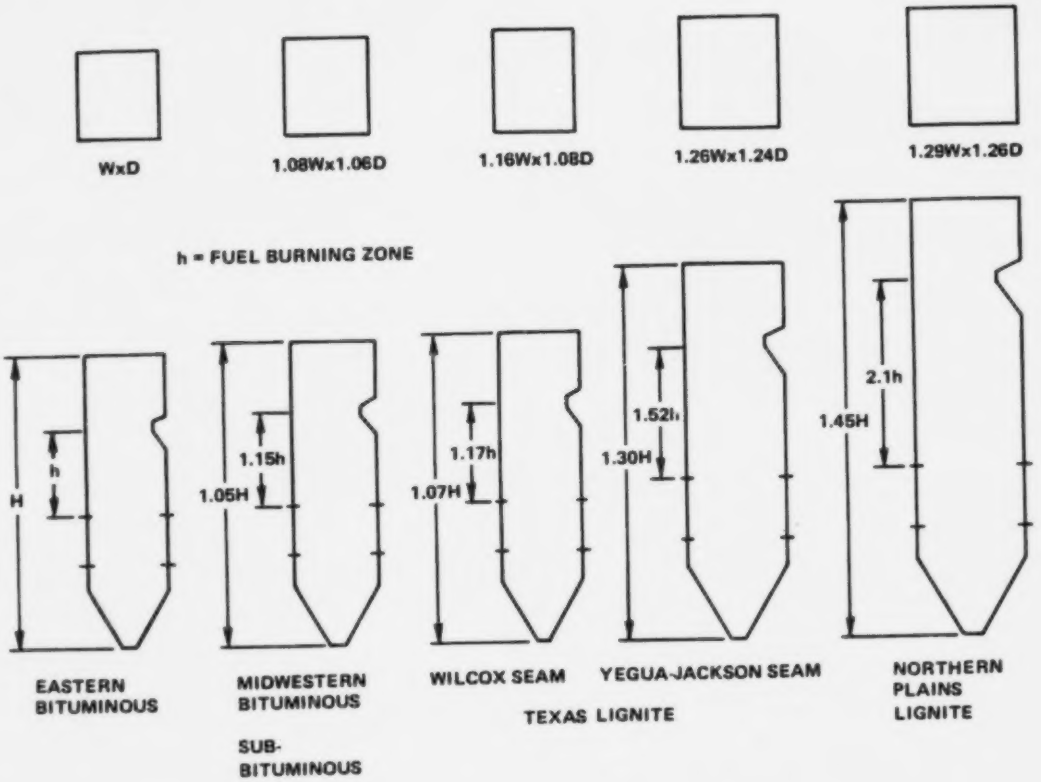

WxD
1.08Wx1.06D
1.16Wx1.08D
1.26Wx1.24D
1.29Wx1.26D
h = FUEL BURNING ZONE
H
h
1.05H
1.15h
1.07H
1.17h
1.30H
1.52h
1.45H
2.1h
EASTERN BITUMINOUS
MIDWESTERN BITUMINOUS
SUB-BITUMINOUS
WILCOX SEAM
YEGUA-JACKSON SEAM
TEXAS LIGNITE
NORTHERN PLAINS LIGNITE

Figure 55. Effect of coal rank on furnace sizing (constant fuel fired).

Figures 56 and 57 show unit plan area and unit height, respectively, versus unit size (MWe) as a function of coal type. To illustrate the above discussion, it can be seen that, at a given unit size, furnaces designed to burn lignite have larger plan areas and are taller than units designed to burn Midwestern bituminous, a higher rank fuel.

8.4 Net Heat Input per Plan Area and Volumetric Heat Release Rate

Among the more important parameters in coal-fired furnace design is net heat input per plan area (NHI/PA, Btu/hr-ft^2). Selection of an appropriate value of NHI/PA requires thorough evaluation of all of the fuels expected to be burned in the unit. Depending upon the coal, current large pulverized coal units are designed for the NHI/PA range of 1.4 to 2.0 x 10^6 Btu/hr-ft^2.

Figures 58 and 59 show NHI/PA and volumetric heat release rate (Btu/hr-ft^3), respectively, versus contract year for the 179 domestic coal-fired units sold between 1960 and 1980. It can be seen that both parameters tend to decrease during the 1965-1975 time period, reflecting the trend toward larger furnaces. This trend eased during the late 1970's, with both parameters somewhat flatter. The effect of coal rank on these parameters (larger plan area with poorer coal) can also be seen.

8.5 Number of Burners per Unit

Essentially all utility boilers designed by Combustion Engineering, Inc. are equipped with tangentially fired burners. With tangential firing, fuel and air are introduced into the furnace at multiple elevations from the four furnace corners. The fuel and air (primary) to each burner elevation are supplied by one mill (or pulverizer). The number of mills depends on the unit size and the type of fuel burned. For units over 100 MWe, a minimum of four mills are generally required.

Figures 60 through 63 show the number of burners per furnace versus unit size (MWe) for Midwestern bituminous, Easter bituminous, sub-bituminous, and lignite coals, respectively. Generally, the number of burners increases with

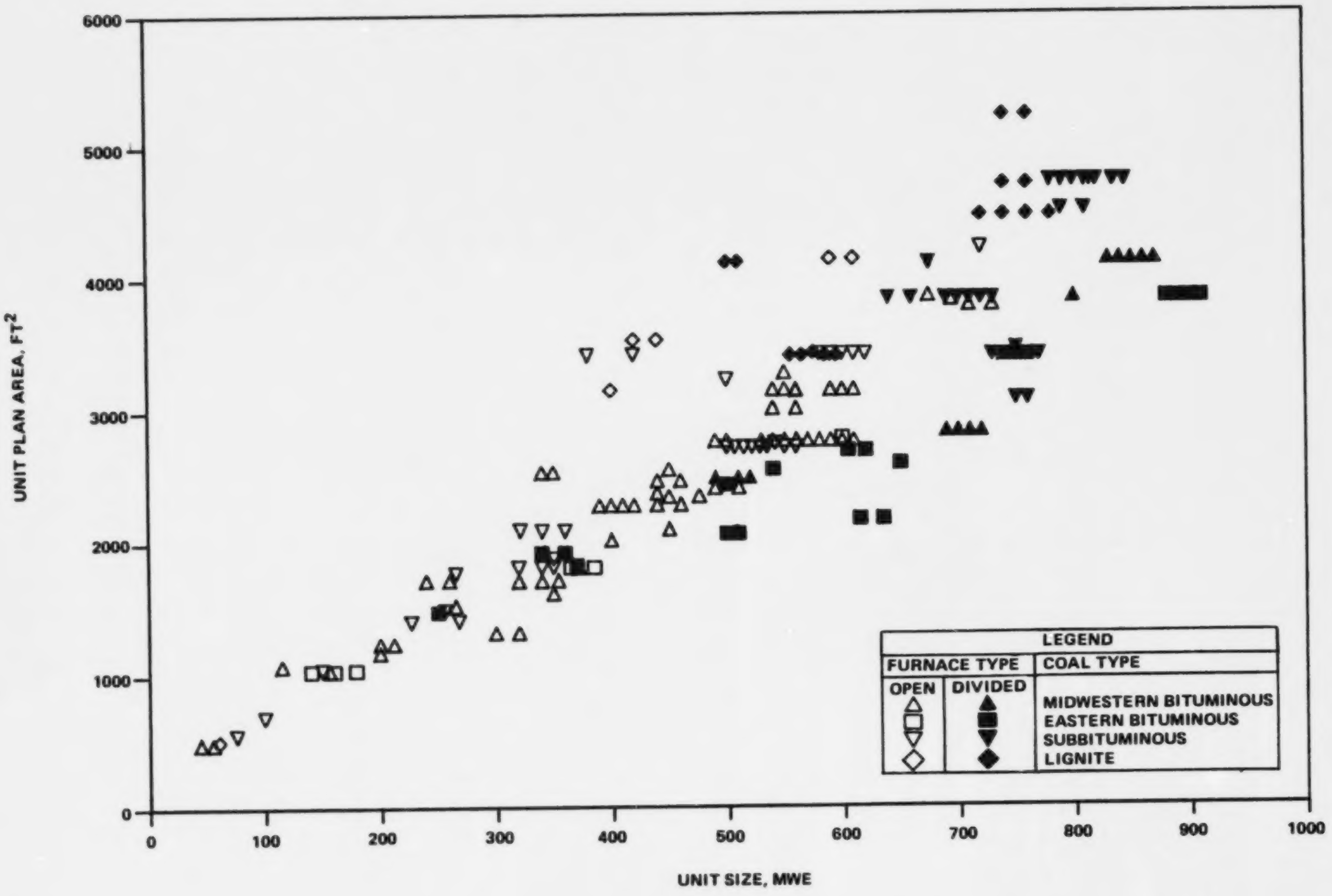


Figure 56. Unit plan area versus unit size for C-E domestic utility coal-fired boilers (1960-1980).

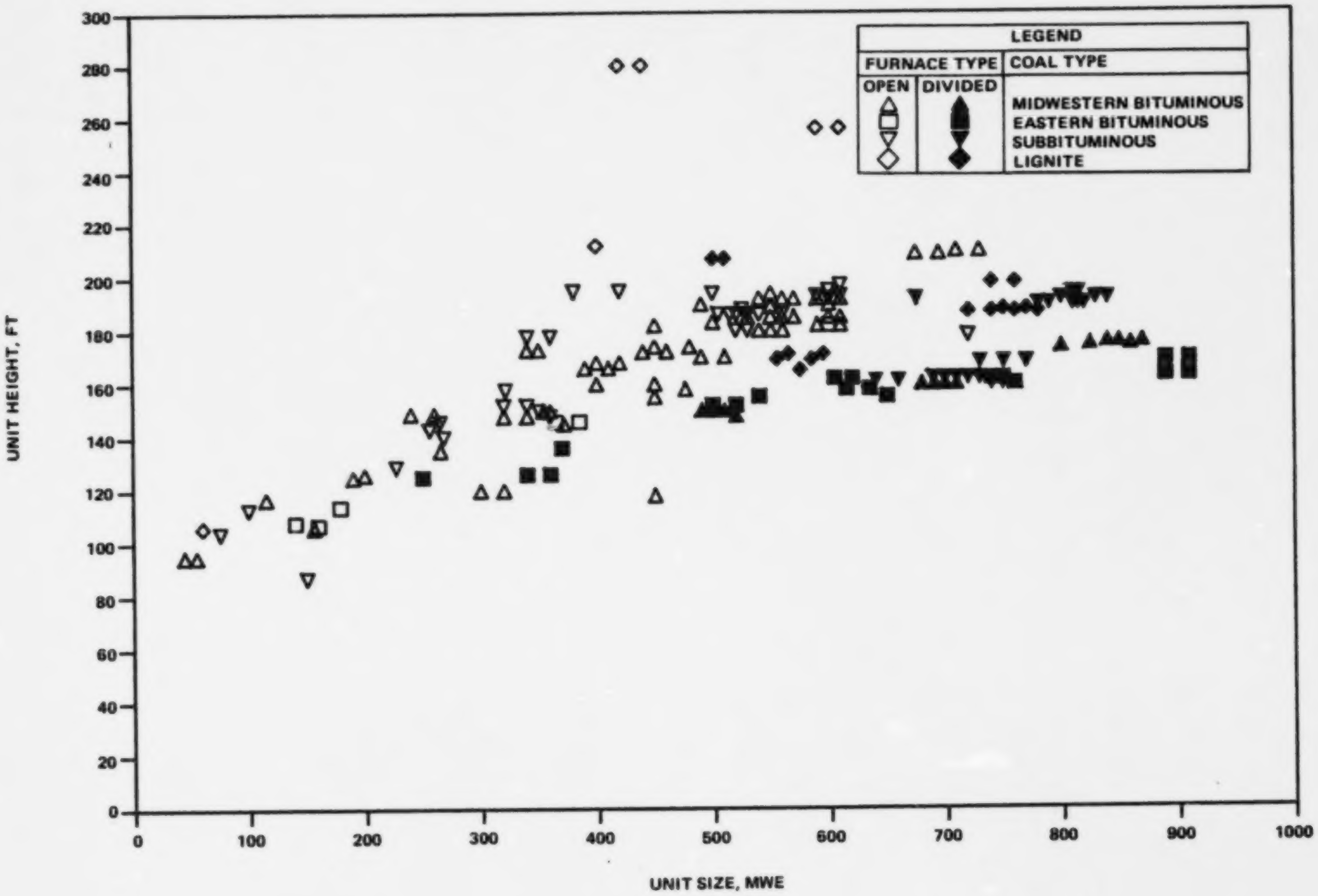


Figure 57. Unit height versus unit size for C-E domestic utility coal-fired boilers (1960-1980).

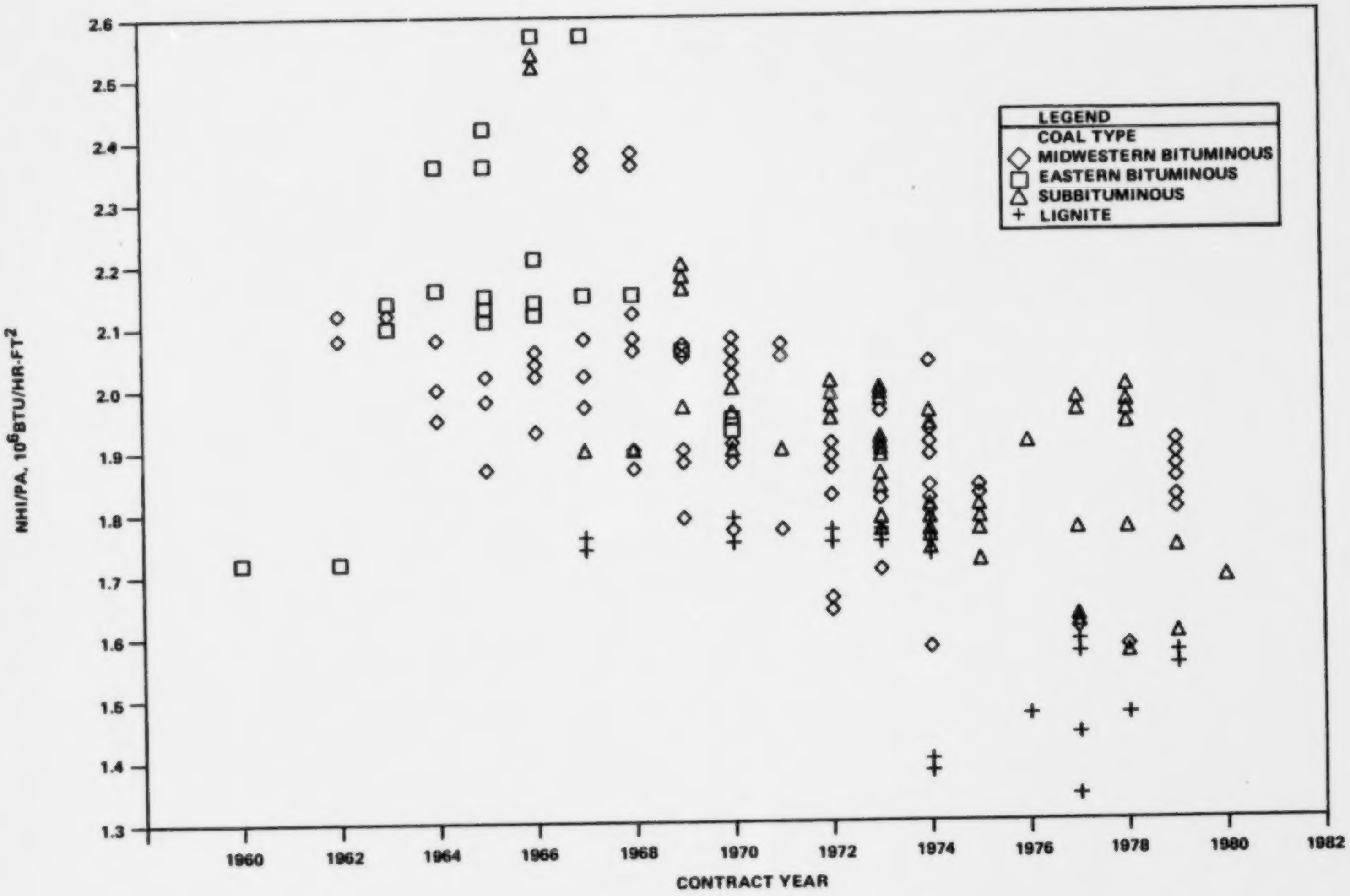


Figure 58. Net heat input per plan area versus contract year for C-E domestic utility coal-fired boilers (1960-1980).

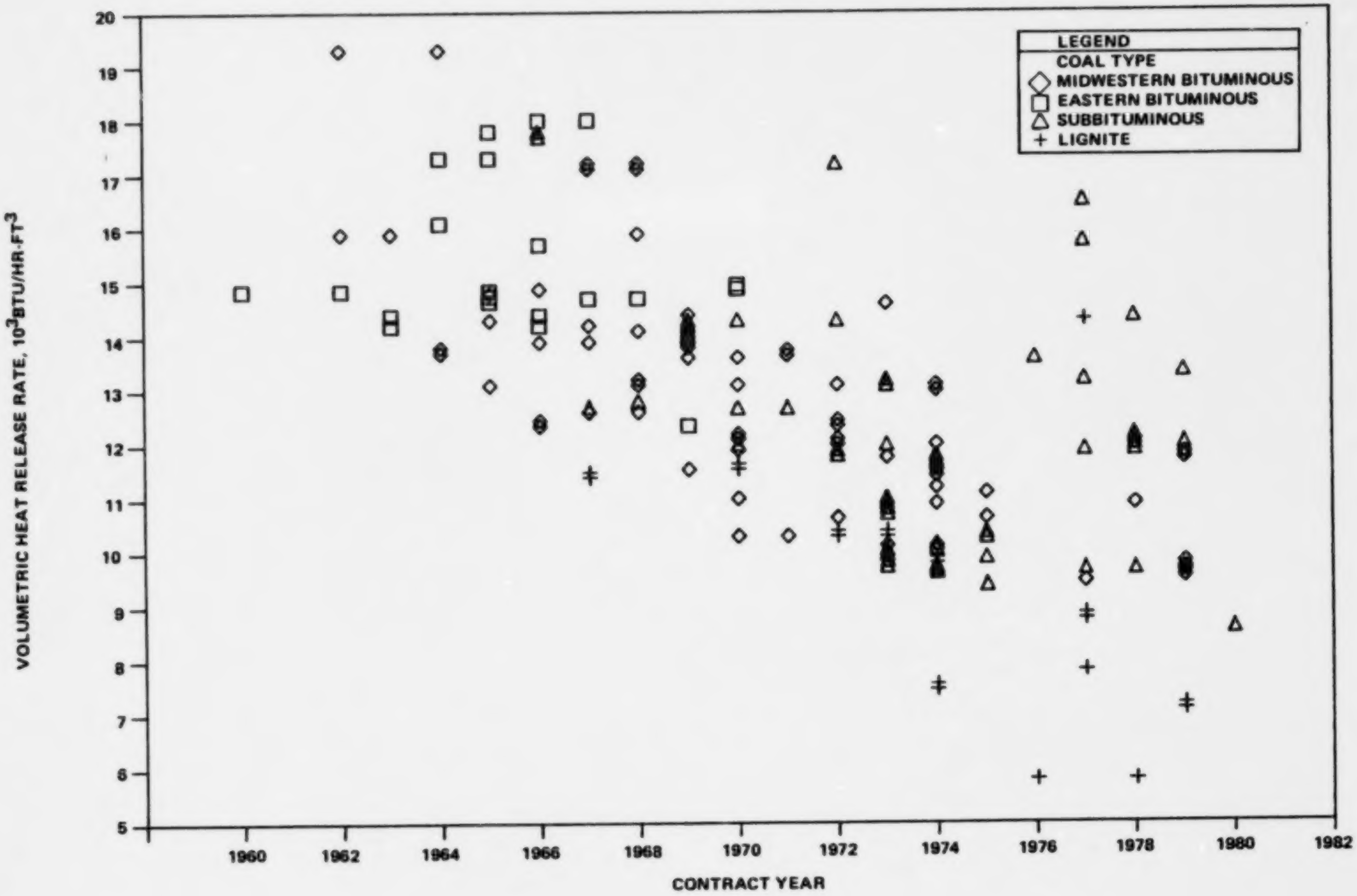


Figure 59. Volumetric heat release rate versus contract year for C-E domestic utility coal-fired boilers (1960-1980).

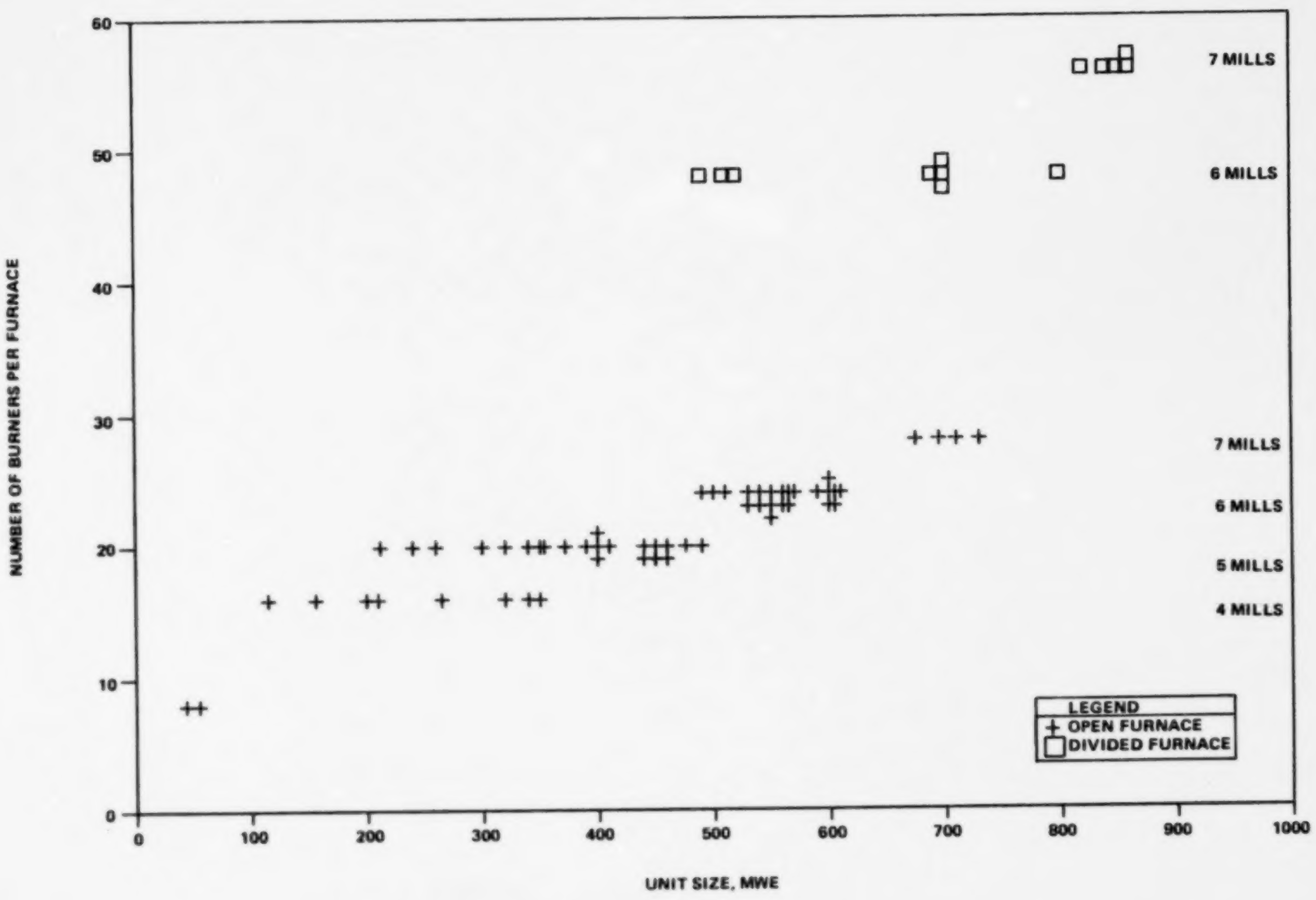


Figure 60. Number of burners per furnace versus unit size for C-E units burning Midwestern bituminous coal (1960-1980).

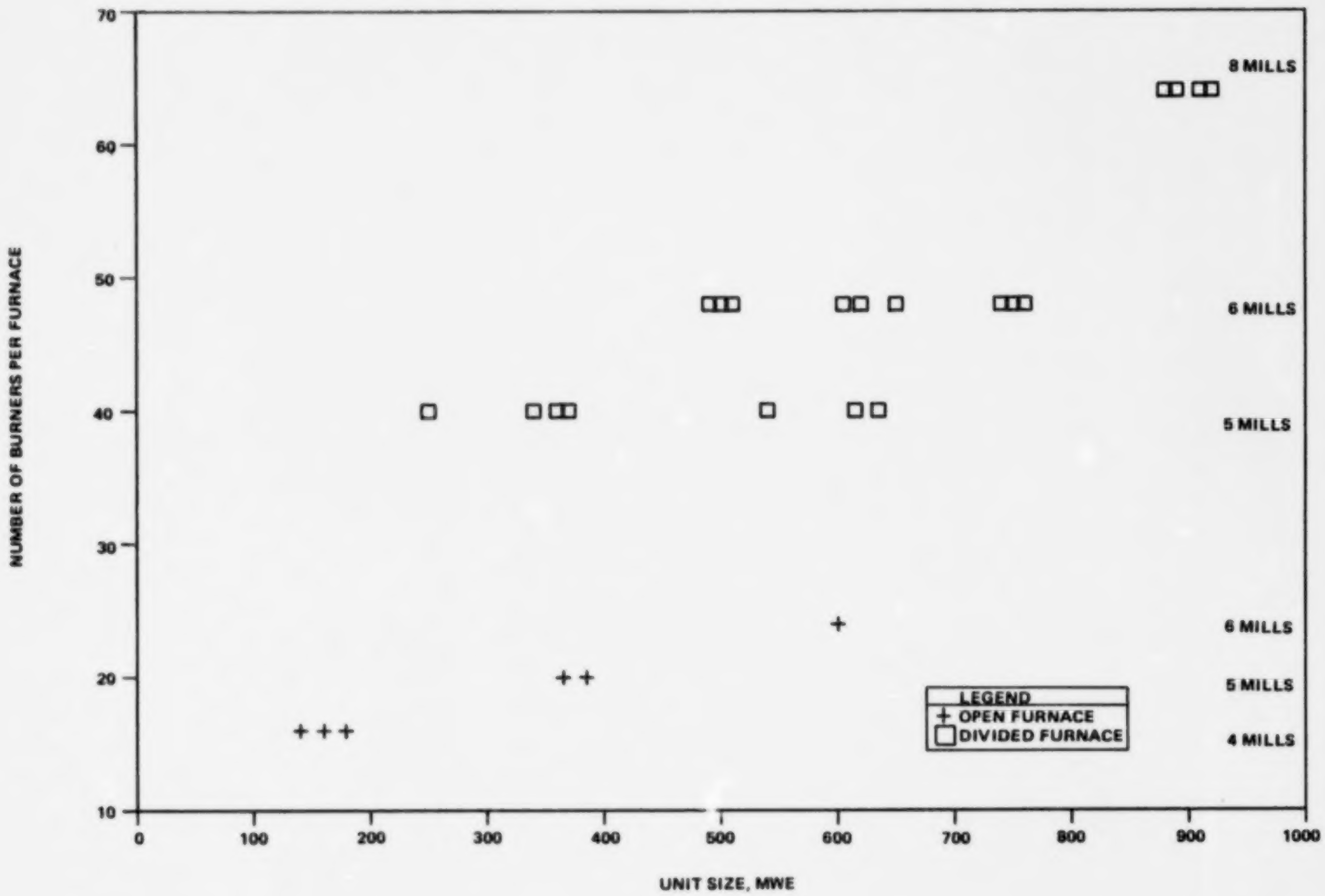


Figure 61. Number of burners per furnace versus unit size for C-E units burning Eastern bituminous coal.

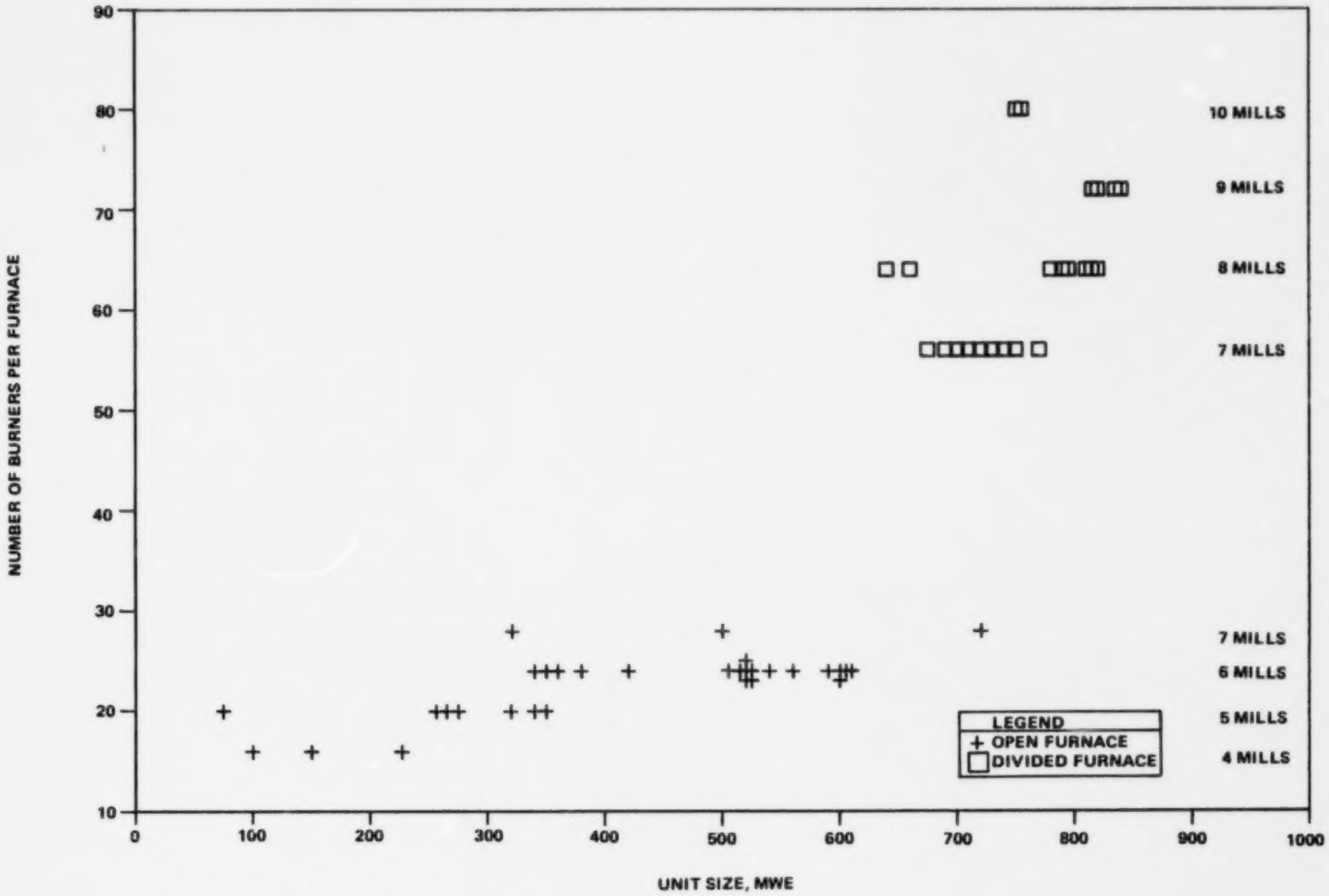


Figure 62. Number of burners per furnace versus unit size for C-E units burning subbituminous coal (1960-1980).

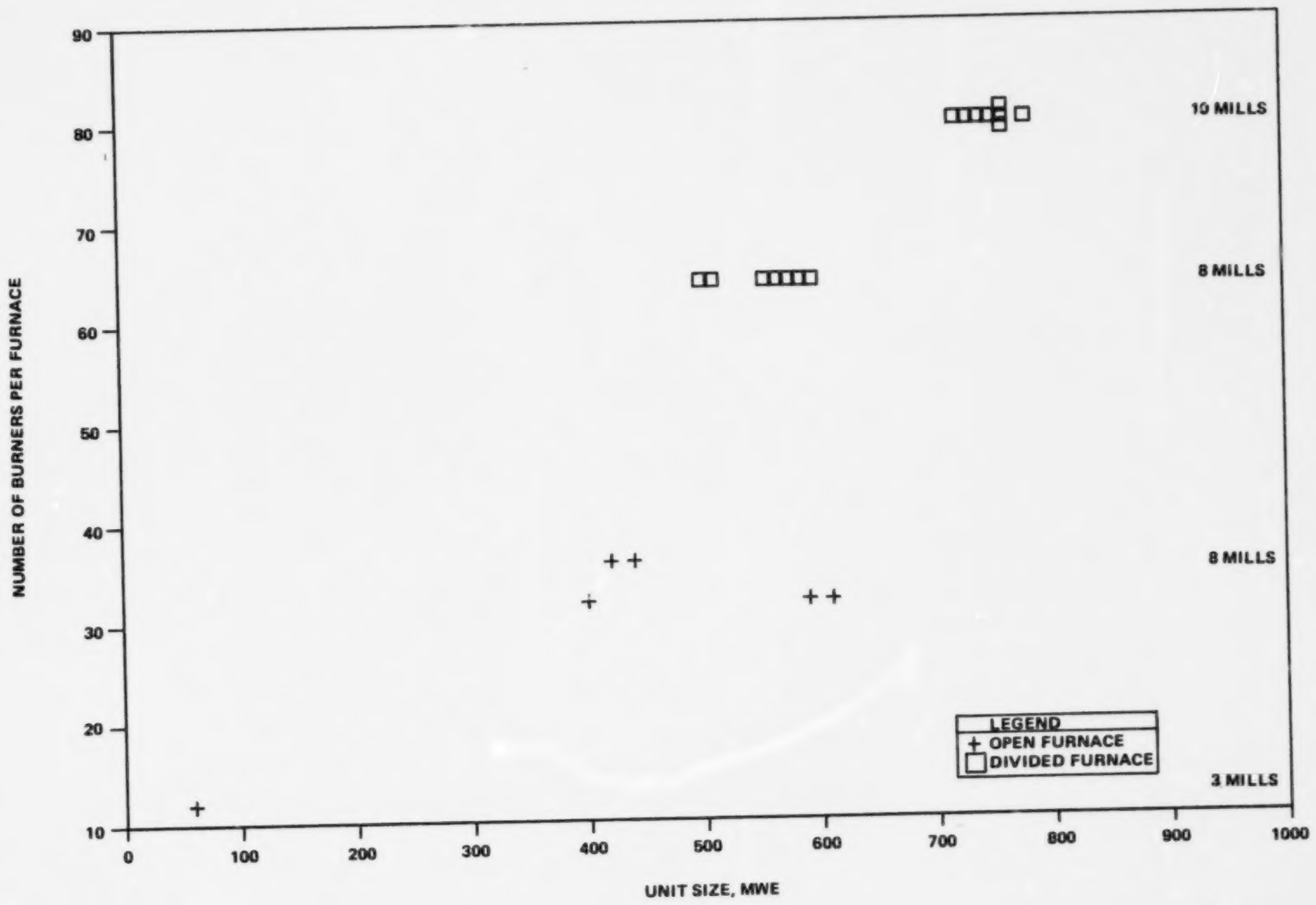


Figure 63. Number of burners per furnace versus unit size for C-E units burning lignite (1960-1980).

unit size. This number also depends on the number and size of the mills. For example, a 200 MWe unit may have the same number of burners as a 500 MWe unit, but, because larger mills are used for the larger unit, fewer mills are required. The selection of mill size and the number of mills depends on fuel properties such as grindability, moisture content, and heating value. For low grade coals, more and larger mills are required as shown in Figure 63. Divided furnaces have twice as many burners for a given number of mills, since each mill feeds burners in eight corners.

8.6 Heat Input per Burner

The maximum allowable heat input per burner (HIPB, Btu/hr/burner) depends upon the unit plan area and the coal ash properties. The actual HIPB can vary from 50 million Btu/hr to a maximum of 300 million Btu/hr.

Figures 64 through 67 show HIPB versus unit size (MWe) for Miswestern bituminous, Eastern bituminous, subbituminous, and lignite coals, respectively. It can be seen that, in general, HIPB increases with unit size. However, these figures also show that HIPB is fuel dependent; for low grade lignite, HIPB is lower than for the higher grade coals. It can also be seen that, for the same unit size, HIPB for divided furnaces is approximately half that for open furnaces. This is due to the divided furnaces having twice as many corners, thus requiring twice as many burners.

8.7 Number of Sootblowers

The sootblowing system is one of the most important auxiliaries on a coal-fired unit. The purpose of this system is to maintain proper heat transfer by removing accumulated deposits to prevent convective pass plugging. Excessive buildup of deposits can result in load reductions. Higher local velocities due to these flow constrictions can also result in increased tube erosion.

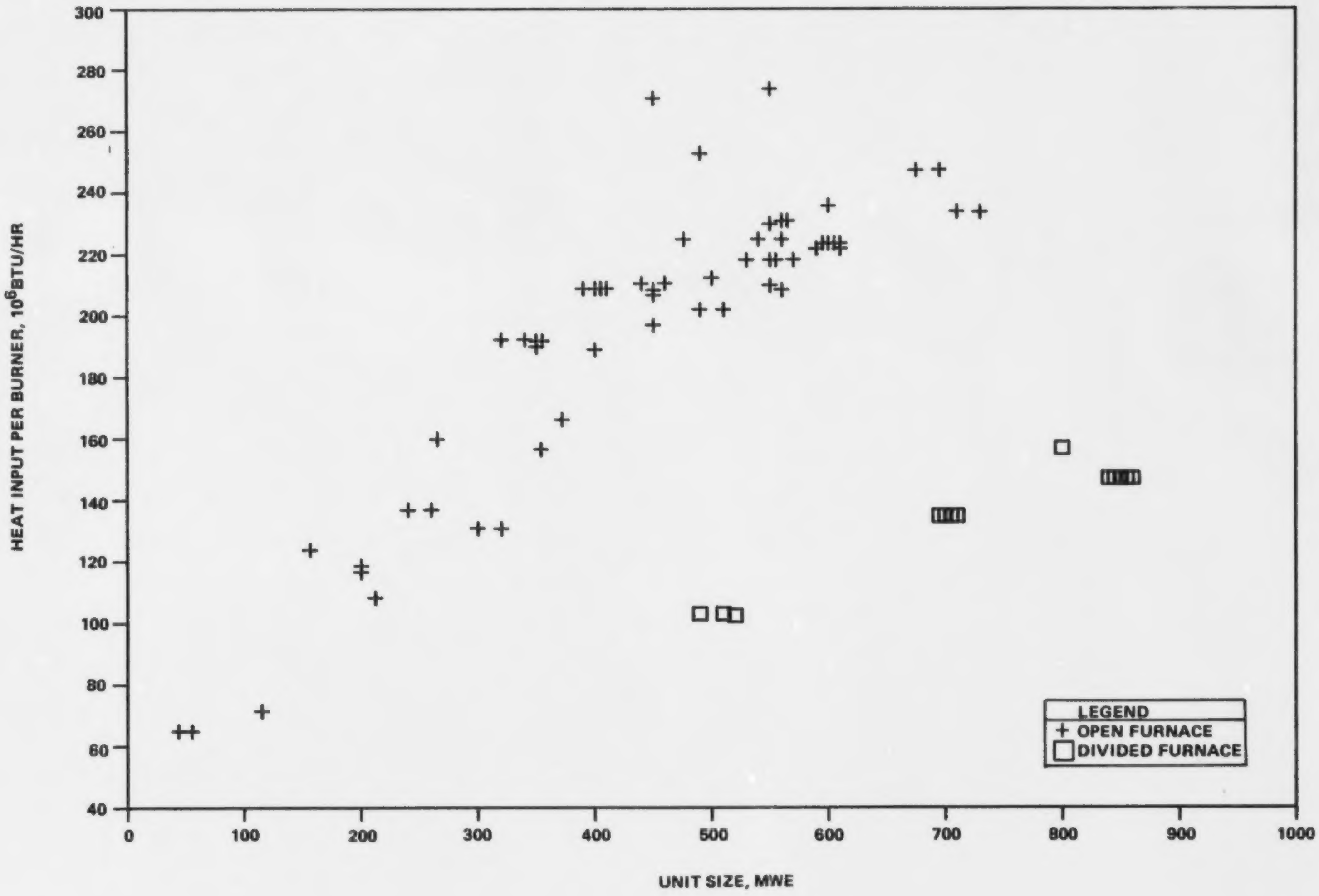


Figure 64. Heat input per burner versus unit size for C-E units burning Midwestern bituminous coal (1960-1980).

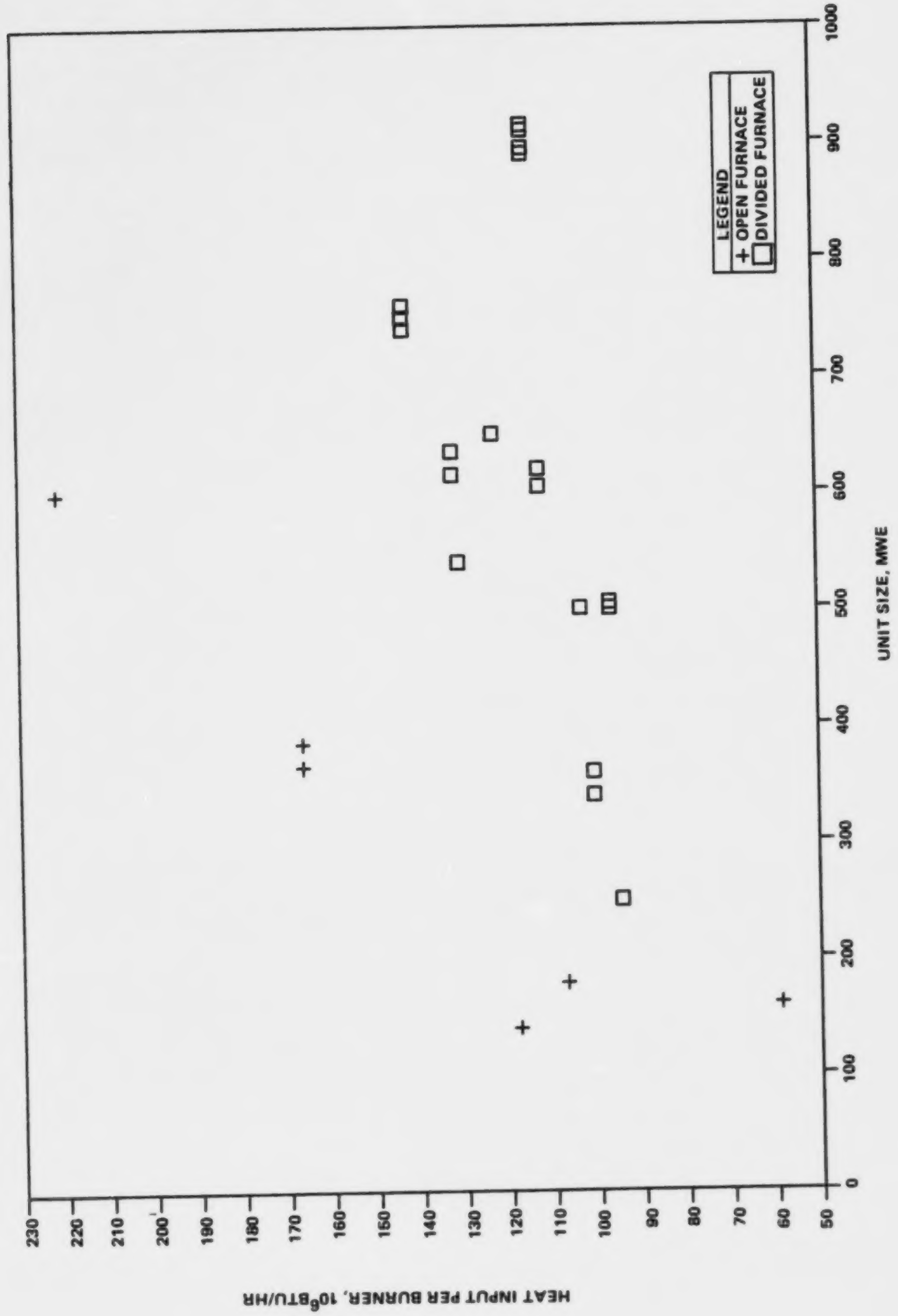


Figure 65. Heat input per burner versus unit size for C-E units burning Eastern bituminous coal (1960-1980).

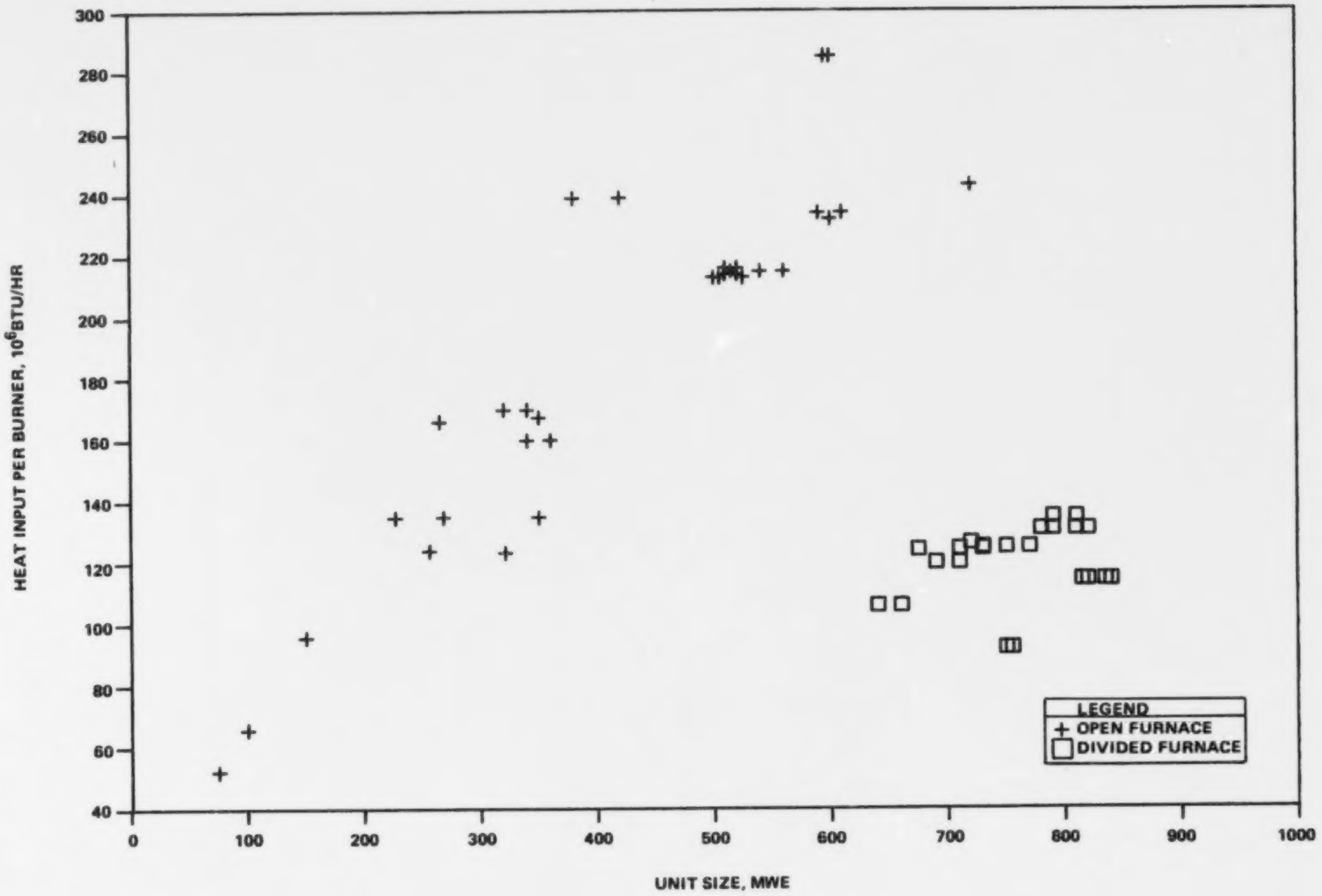


Figure 66. Heat input per burner versus unit size for C-E units burning subbituminous coal (1960-1980).

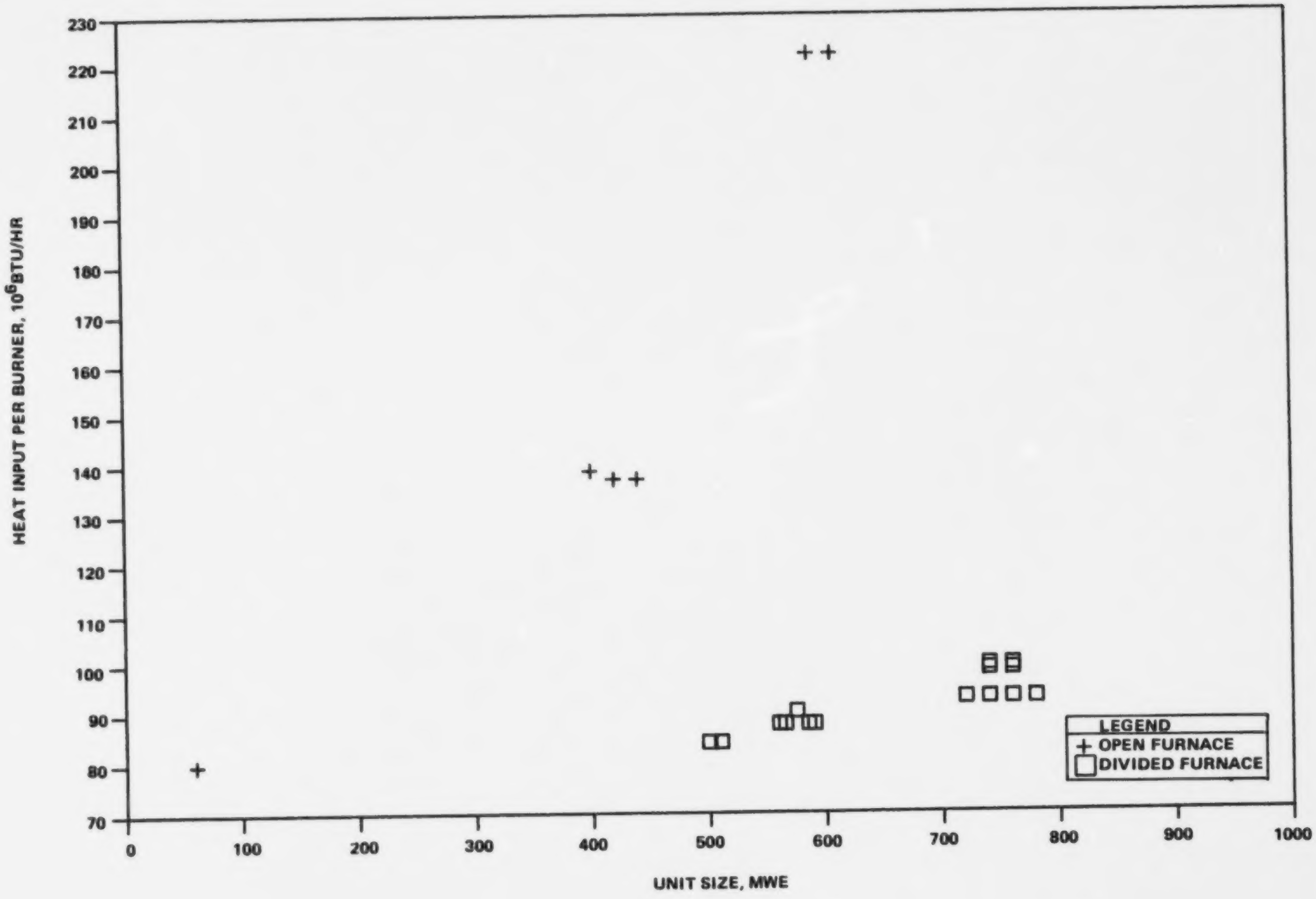


Figure 67. Heat input per burner versus unit size for C-E units burning lignite (1960-1980).

Two types of sootblowers are used in utility boilers. Furnace wall deposits are removed by wall blowers. These short single-nozzle retractable blowers penetrate the furnace walls one or two inches and rotate through 360°. Convective section deposits are removed by retractable blowers. Long fully retractable lances penetrate the cavities between the convective sections. Two 180° opposed cleaning nozzles at the tip are normally used. While the lance traverses the boiler, it rotates, forming a helical blowing pattern which effectively cleans the tubes and unplugs the spaces between tubes. The cleaning medium for both types of sootblower can be either superheated steam or compressed air. The utility normally chooses the medium based upon plant economics.

The number of sootblowers required for a particular unit depends on the rate of deposit and the characteristics of the deposited material. The type of deposit in the radiant and convective sections of the boiler can vary from very hard tenacious slag to a dry powdery coating. The most important coal-ash properties affecting the severity and rate of deposit are the ash softening and fluid temperatures, percent ash in the coal, and percent sodium in the coal ash. Coals with high ash contents, high ash fusion temperature, and high percentages of sodium in the ash present the most difficult cleaning problems.

Figures 68 through 71 show the number of wall sootblowers versus unit size (MWe) for Midwestern bituminous, Eastern bituminous, subbituminous, and lignite coals, respectively. It can be seen that the number of sootblowers required is primarily dependent on coal rank, with units burning lignite requiring the greatest number of wall blowers at a given unit size. The substantial amount of scatter for a given coal shows the effect of variation in the properties of that coal.

Figures 72 through 75 shows the number of retractable sootblowers versus unit size for the four coal types. Again, the wide variation in the number of sootblowers required shows that coal rank and variations in coal properties within a rank are the predominant factors affecting the number of sootblowers required. The figures also show that the number of retractable sootblowers

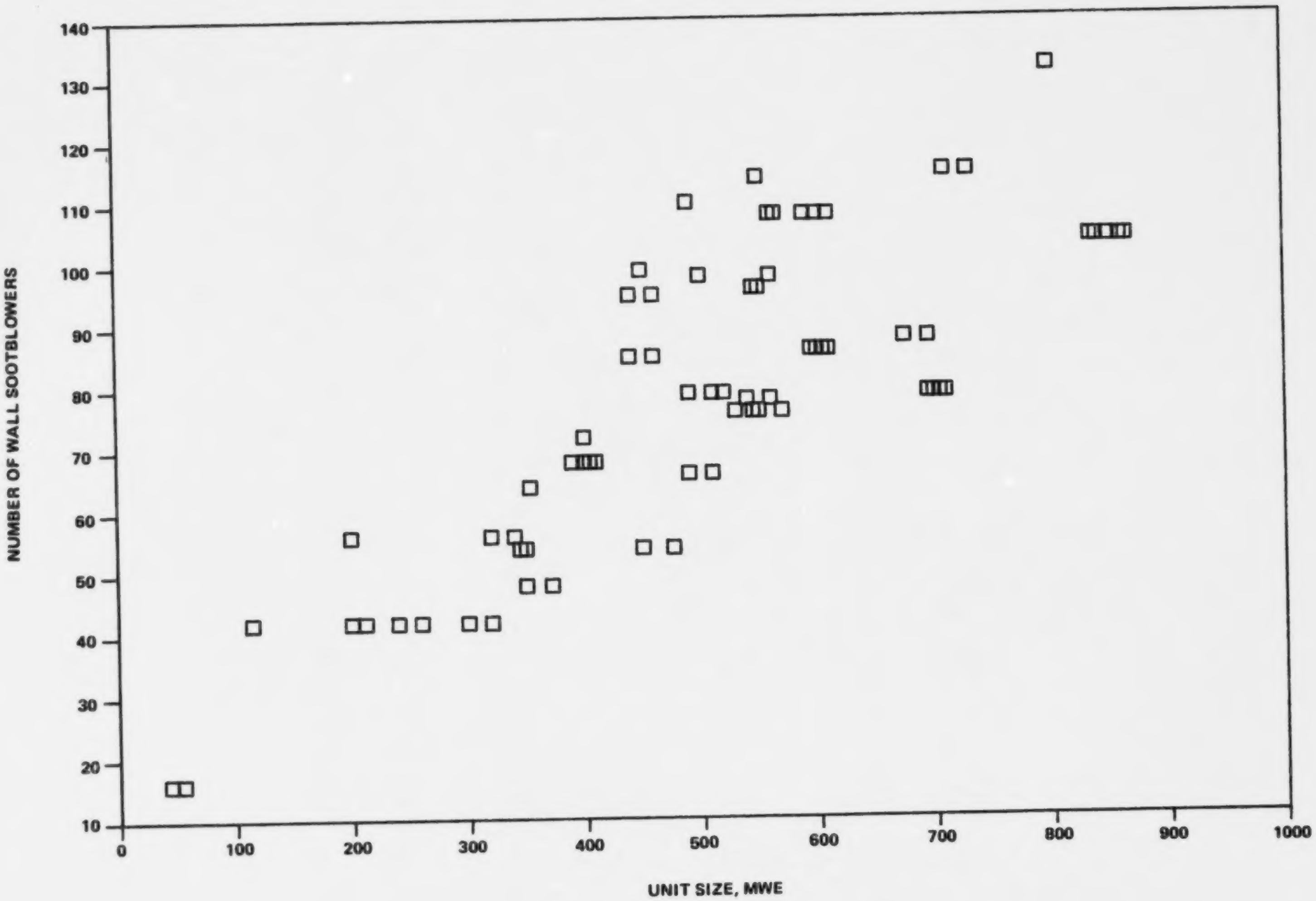


Figure 68. Number of wall sootblowers versus unit size for C-E units burning Midwestern bituminous coal (1960-1980).

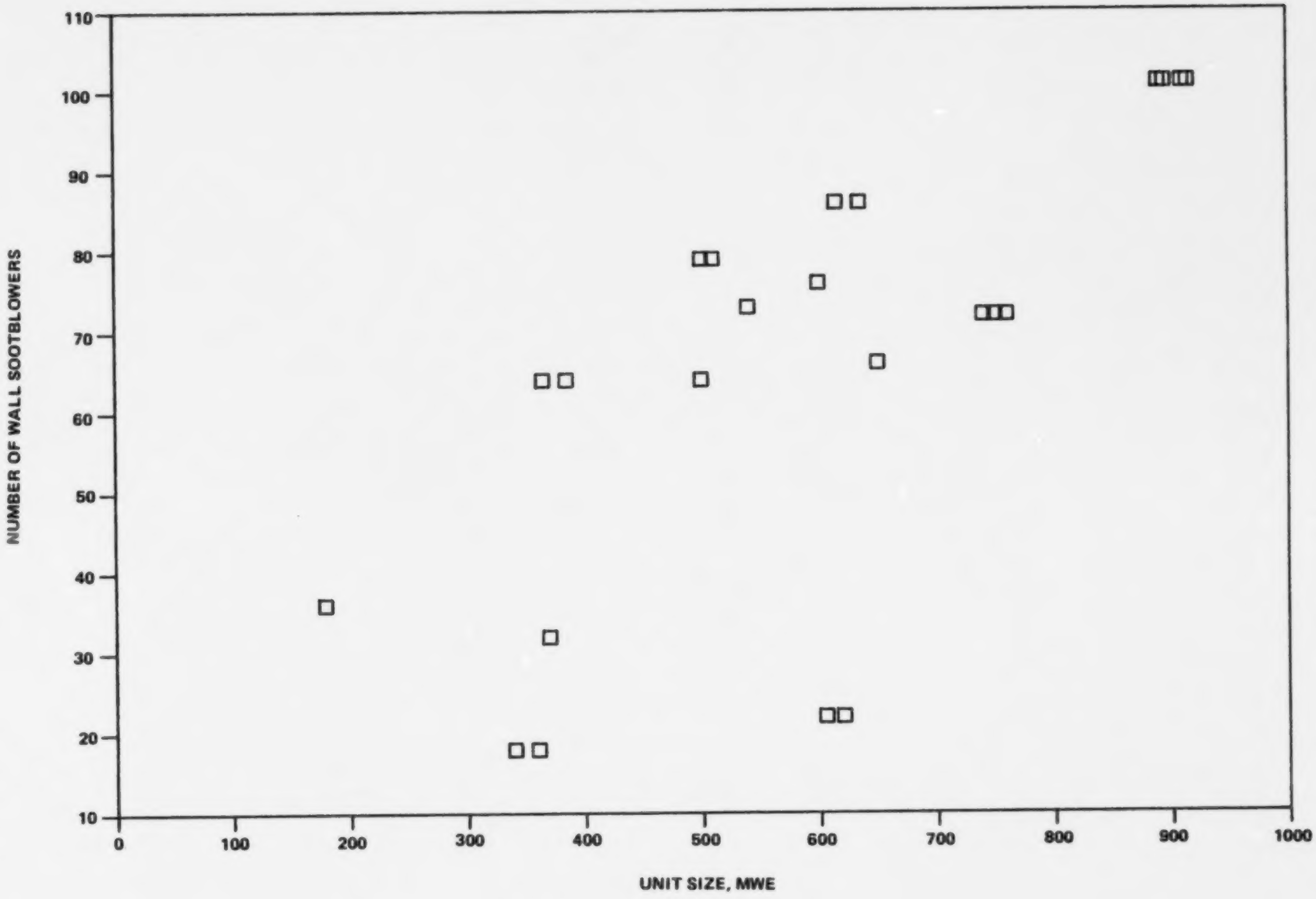


Figure 69. Number of wall sootblowers versus unit size for C-E units burning Eastern bituminous coal (1960-1980).

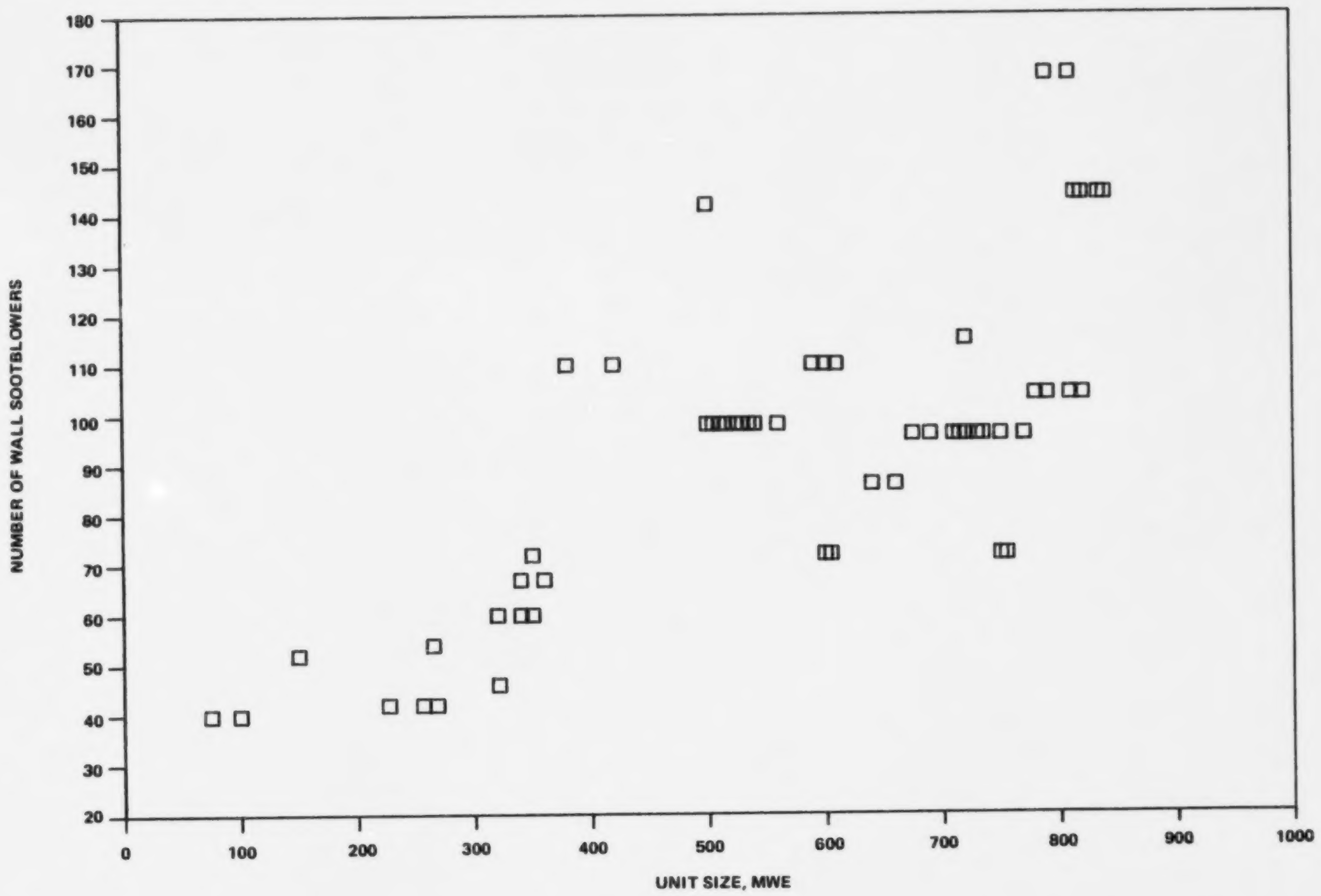


Figure 70. Number of wall sootblowers versus unit size for C-E units burning subbituminous coal (1960-1980).

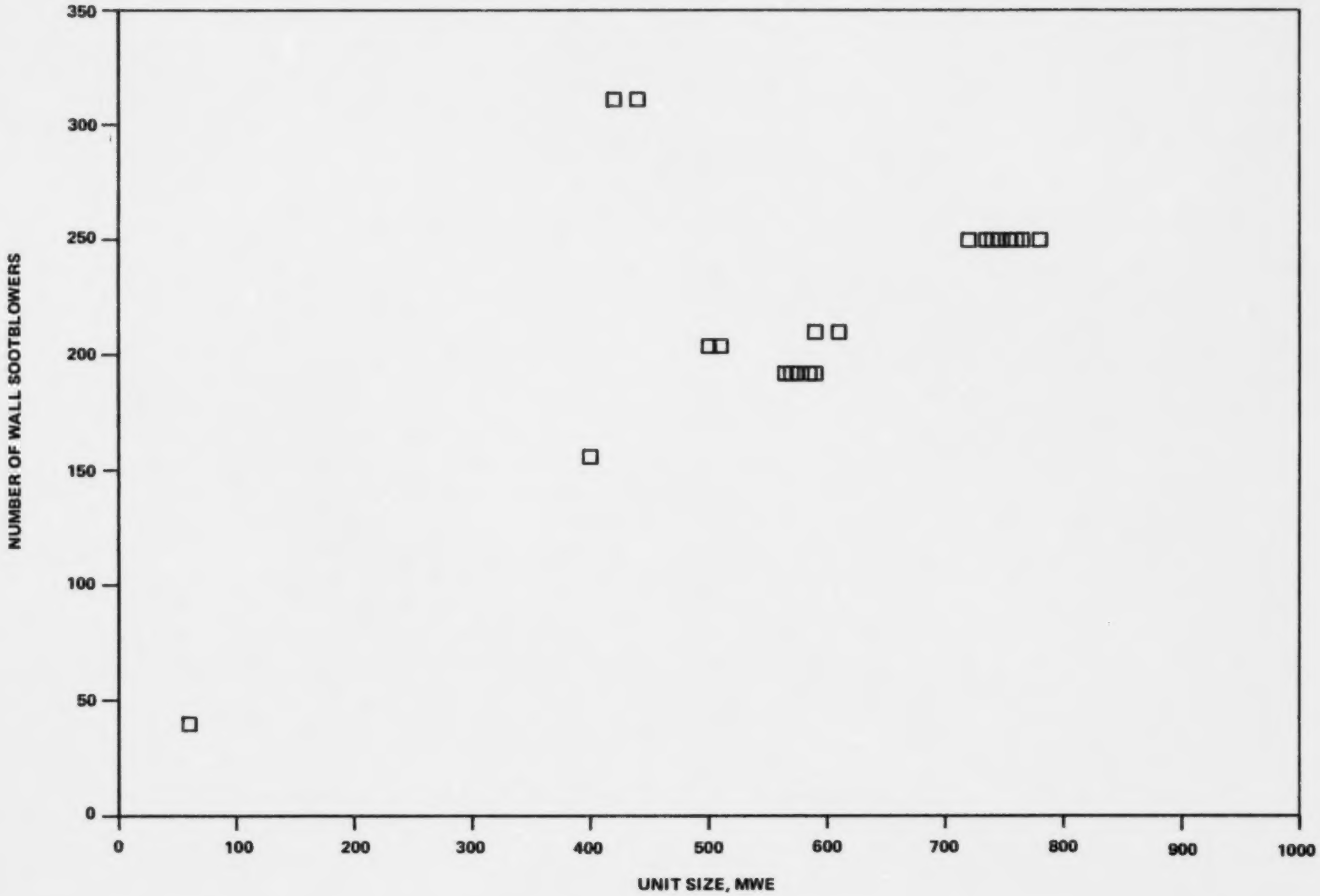


Figure 71. Number of wall sootblowers versus unit size for C-E units burning lignite (1960-1980).

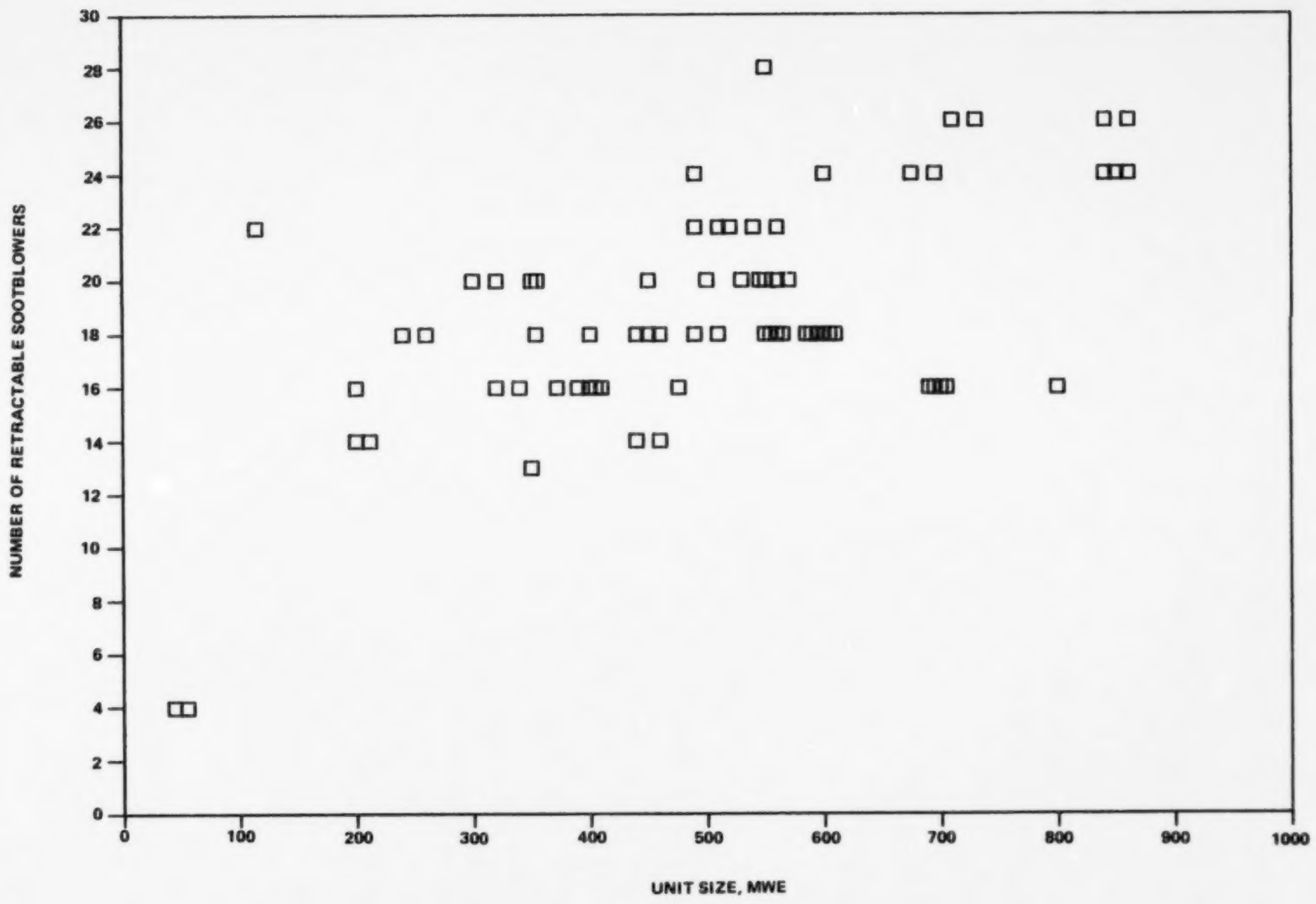


Figure 72. Number of retractable sootblowers versus unit size for C-E units burning Midwestern bituminous coal (1960-1980).

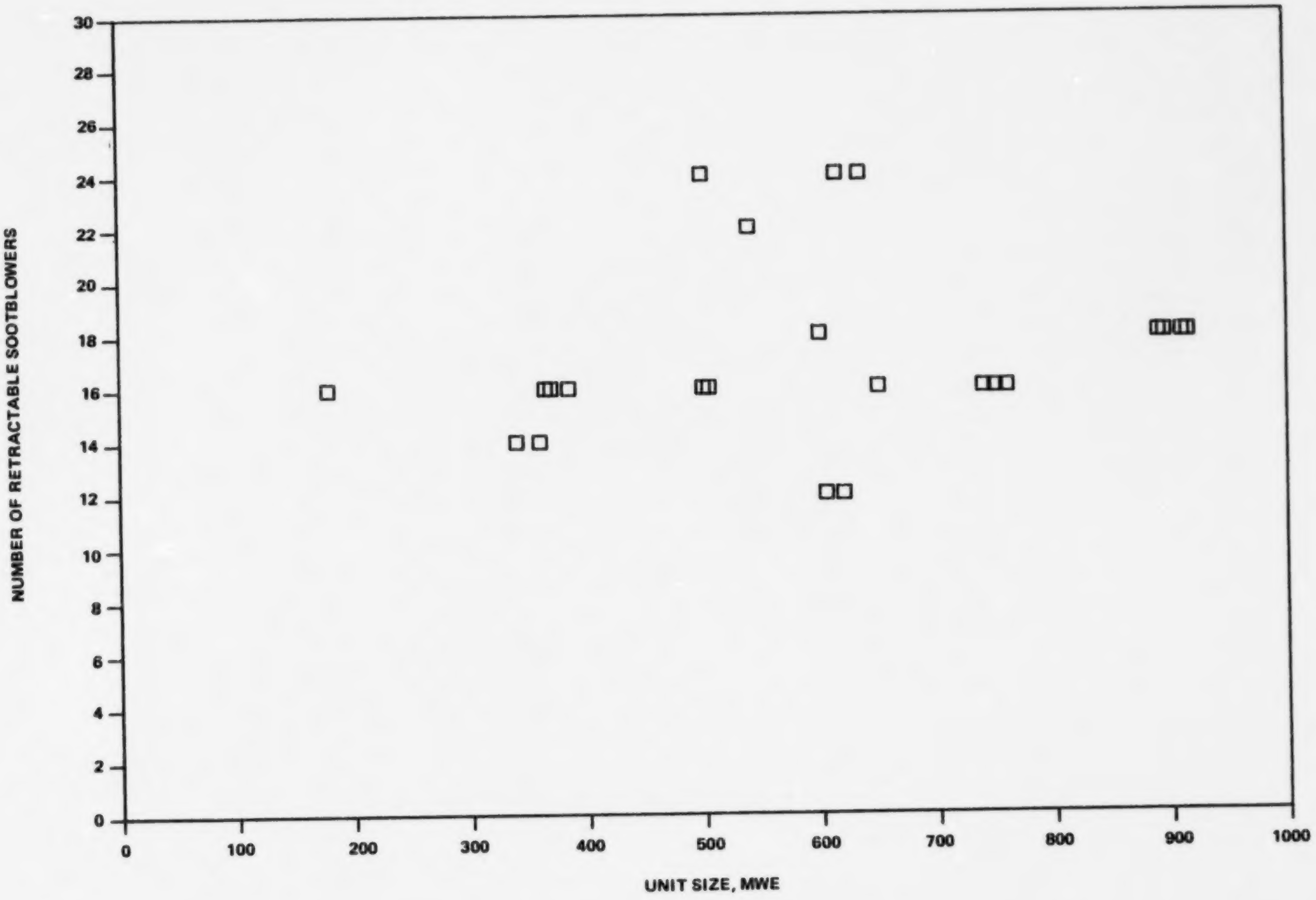


Figure 73. Number of retractable sootblowers versus unit size for C-E units burning Eastern bituminous coal (1960-1980).

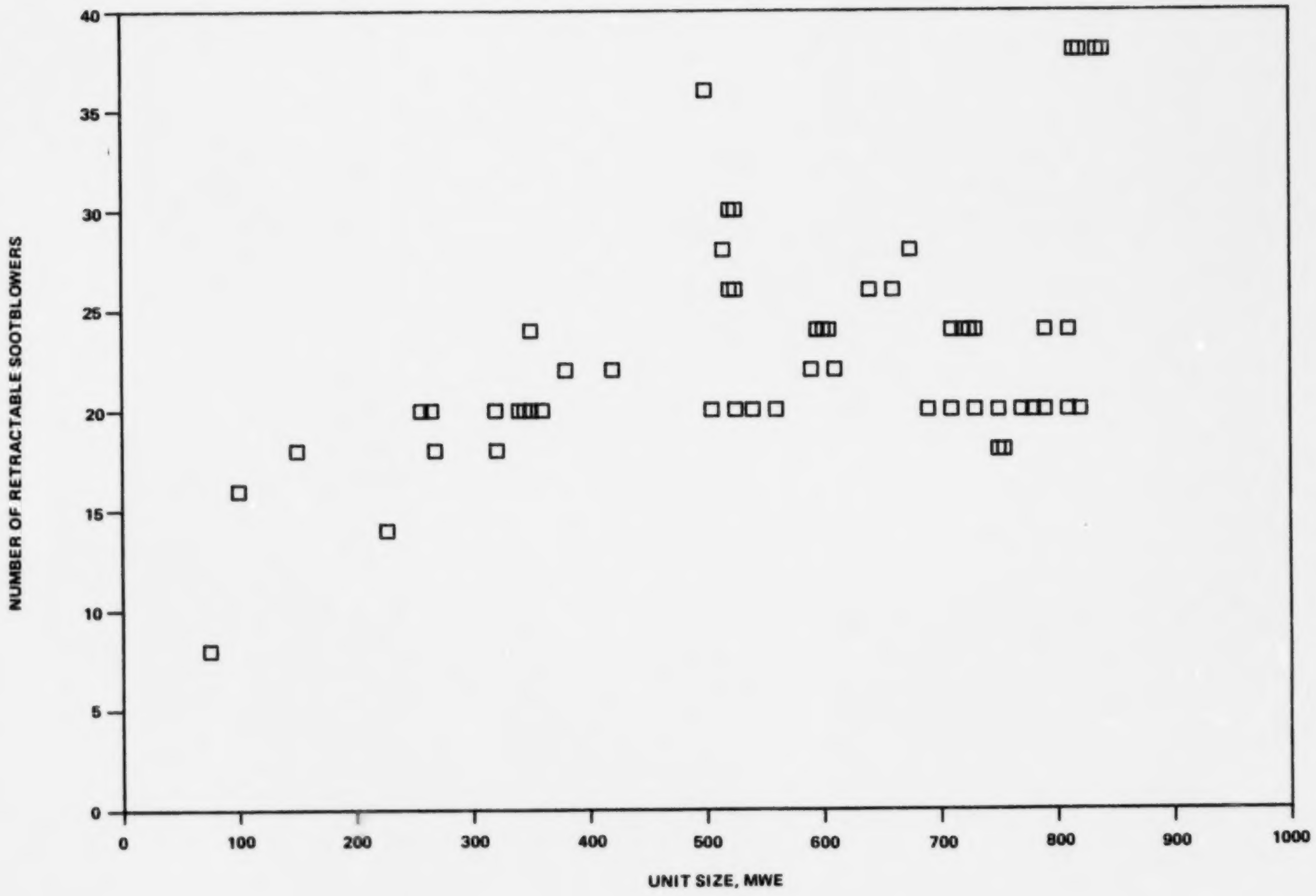


Figure 74. Number of retractable sootblowers versus unit size for C-E units burning subbituminous coal (1960-1980).

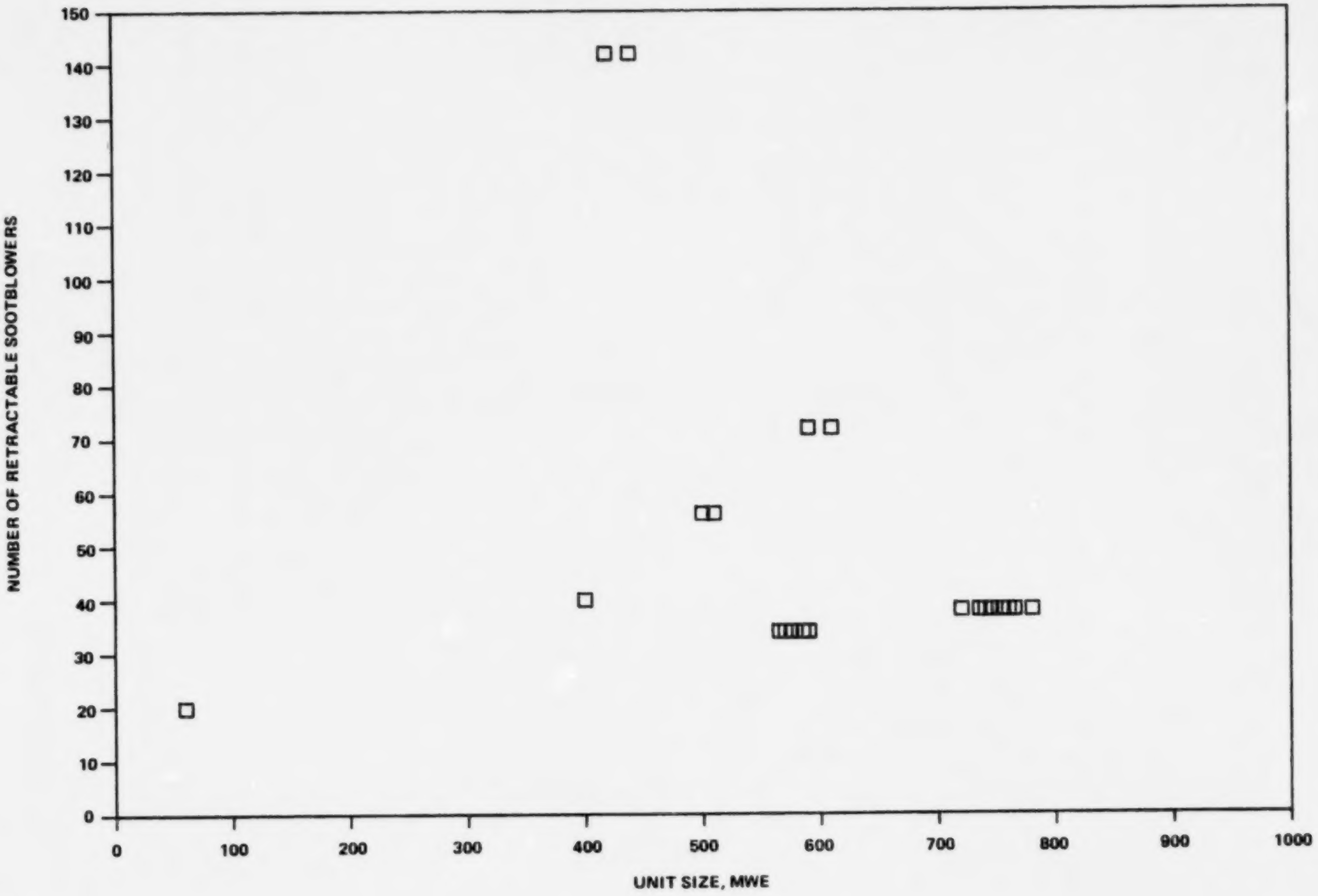


Figure 75. Number of retractable sootblowers versus unit size for C-E units burning lignite (1960-1980).

required for a given unit is almost independent of unit size. This may be due to the fact that the convective sections of larger units can be cleaned by using longer lances rather than with a greater number of sootblowers.

8.8 Air Quality Control Systems

Three classes of emissions from utility boilers are judged to be significant from an air quality standpoint: particulate matter, sulfur oxides, and nitrogen oxides. Historically, particulate matter has received the greatest attention because it is easily seen and is often labeled a public nuisance. The concern about sulfur oxides comes from its possible health effects and its potential for damaging vegetation due to acidic precipitation. Oxides of nitrogen are significant because they participate in complex chemical reactions that lead to formation of photochemical smog in the atmosphere. The effects of the smog, both with regard to loss of visibility and as an eye and respiratory irritant, are well known.

8.8.1 Particulate Removal

Particulate emission from coal-fired boilers varies considerably, depending upon the type of firing employed and the ash content of the coal. A pulverized coal-fired unit can normally expect to have 60% to 80% of the coal ash leaving the furnace with the flue gas. The remaining ash leaves through the boiler bottom-ash removal system.

A wide variety of equipment is available for control of particulate emission from utility boilers. Table 11 lists types of systems which have been installed on domestic C-E coal-fired units. (Detailed descriptions of these systems is beyond the scope of this report). With the exception of the mechanical collector, these systems have the potential for achieving a removal efficiency of over 99%. It can be seen that 93% of the domestic C-E coal-fired units incorporate some form of particulate control device, with the electrostatic precipitator the overwhelming choice.

TABLE 11. DISTRIBUTION OF AIR QUALITY CONTROL SYSTEMS ON C-E COAL-FIRED UTILITY UNITS (1960-1980)

	No. of Units	% of Total
Particulate control		
Electrostatic precipitator	146	
Wet scrubber	7	
Fabric filter	12	
Mechanical cylone collector	1	
Total	166	93
Flue gas desulfurization		
Limestone	22	
Lime	11	
Lime/limestone	4	
Lime/alkali flyash	9	
Double alkali	1	
Sodium carbonate	1	
Lime/spray drying	4	
Total	52	29
NO_x control		
Overfire air	91	
Total	91	51

8.8.2 Flue Gas Desulfurization (FGD)

As fuel burns, most of the sulfur contained in the fuel is converted to sulfur oxides (SO_2 and SO_3). The quantity of sulfur in coal varies widely, from 0.4% to over 5%. In general, at least 90% of this sulfur will oxidize, constituting from 0.1 to 0.25 percent by volume of the combustion products.

There are many processes available for removing sulfur oxides from flue gas (again, detailed descriptions of these processes are beyond the scope of this report). Table 11 shows the various FGD systems which have been installed on C-E coal-fired boilers. It can be seen that only about 29% of the 179 units have some form of flue gas desulfurization. Approximately half of this group uses systems based on the limestone scrubbing process. Only one of these units was ordered before 1970.

8.8.3 NO_x Control

During the combustion process, nitrogen oxides (NO_x) are formed in the high temperature region in and around the flame zone. The NO_x which is formed comes from both the nitrogen in the air (thermal NO_x) and the nitrogen in the coal (fuel NO_x). The rate of NO_x formation is influenced by both the temperature and the amount of oxygen which is present. Thermal NO_x formation can be controlled by reducing the flame temperature and the amount of excess air. Controlling the amount of oxygen in the combustion region has the most significant effect on fuel NO_x.

Overfire air operation is one method for controlling NO_x which has proven to be both successful and economical on units equipped with tangential burners. This method lowers the flame temperature by burning the fuel in two stages. In this manner, all of the heat is not released at once. This is accomplished by burning the fuel with slightly above stoichiometric air in the burner zone. The remainder of the air is introduced through overfire air ports above the burner zone at which point the rest of the fuel is consumed. Lowering the flame temperature in the burner zone reduces thermal NO_x formation. In addition, the reduction in the amount of oxygen available reduces the fuel NO_x formation. Table 11 shows that 51% of the domestic C-E coal-fired boilers are equipped with overfire air for NO_x control.

SECTION 9

SELECTION RATIONALE FOR CANDIDATE HOST SITE

Successful field demonstration of dry sorbent injection with low-NO_x firing equipment depends strongly upon careful selection of the candidate host site(s). Several design and operating criteria must be met in order to derive valid and meaningful results from the test with the least amount of capital expenditure.

The 179 domestic coal fired units analyzed in Section 8 (Historical and Projected Design Trends) were reviewed to identify likely candidates for retrofitting LIMB process equipment for SO_2 removal. Several criteria were evaluated in reducing the number of units for consideration.

9.1 Primary Selection Process

Three retrofit units were required to be selected for economic and process feasibility studies in the 200, 400, and 600 MWe size ranges, with one of these units having been selected as the candidate host site. Thus, the first step in the selection process was to eliminate units outside the desired size range. Forty-six units above 650 MWe and seven units below 200 MWe were eliminated from consideration. Three groups were established from among the remaining units: 200-270 MWe, 400-476 MWe, and 550-650 MWe.

The next selection criterion considered was coal rank. Because of SO_2 emissions considerations, no units have been designed to burn Eastern bituminous coal since 1970. Ash fusion tests at C-E's Kreisinger Development Laboratory (Section 5) showed that the addition of limestone to Eastern bituminous coal significantly increased the potential for slagging and fouling. Therefore, units which were designed between 1960 and 1970 to burn this type of coal were eliminated from consideration. Units which were designed to burn lignite were also eliminated, since they represent only a small portion of the total number and are not typical of current design practice. At this point in the evaluation and selection process, 53 units remained under consideration. These units are shown in Tables 12 through 14. The units are designated by reference numbers to maintain the anonymity of the utilities involved. As can be seen, several sites have multiple units which are generally identical in design.

9.2 Final Host Site Selection

After reviewing this reduced list of candidate host sites with the EPA Project Officer, at least two units in each size range were selected for further evaluation. Several additional criteria were applied in the evaluation process.

Three units were selected for evaluation of retrofit economics (Sections 10 and 11). They are units No. 10 (560 MWe), 15 (450 MW), and 27 (200 MWe). These units are all located in the central and midwestern regions, are burning Midwestern bituminous coals with at least 3.0% sulfur, are equipped with electrostatic precipitators and are not presently equipped with flue gas desulfurization equipment. All are within short distances of major limestone suppliers. While space is limited at the two larger sites, it is felt that sufficient land is available for both process equipment and materials storage at the 450 MWe site (a limestone pile already exists to feed scrubbers on two other units). For this reason, and because the 200 MWe unit is not felt to be representative of typical C-E utility boiler design (size, open upper furnace), the 450 MWe unit was chosen as the candidate host site for LIMB application.

This steam generator produces 3,145,000 lbs/hr of primary steam and 2,767,000 lbs/hr of reheat steam at the Maximum Continuous Rating condition. The operating pressure is 2621 psig and the turbine throttle pressure is 2521 psig. The unit is of the Controlled Circulation type with a multistage superheater and a two-stage reheater. It is equipped with a continuous finned economizer and two C-E Ljungstrom air heaters.

The unit is not presently equipped with overfire air for NO_x control. Recent operating experience has demonstrated a tendency to slag which could be alleviated by burner modification to an LNCFS configuration.

Particulate emissions are controlled with an ESP located downstream of the air heaters. The precipitator now in use was added to upgrade the removal capability when the unit started to burn the present coal. An older, smaller precipitator is still in place, but is used only as a "knock-out box" as the gas passes through it, collecting a relatively small amount of material.

TABLE 12. CANDIDATE RETROFIT UNITS: 550-650 MWe

			Coal[a]		Gas Cleanup[b]			Type of
Ref. #	Size, MW	No. of Units	Type	% Sulf	FGD	NO_x	Partic	Econ[c]
1	580	2	MW BIT	3.3			ESP	NA
2	600	4	MW BIT	2.9			ESP	CFS
3	600	2	MW BIT	1.8			ESP	CFS
4	600	1	MW BIT	NA			ESP	CFS
5	600	2	SUB C	0.5		OFA	ESP	BARE
6	600	1	SUB A	0.4		OFA	ESP	BARE
7	550	4	MW BIT	0.5	YES	OFA	ESP	CFS
8	560	2	MW BIT	3.2	YES	OFA	ESP	CFS
9	650	2	SUB C	0.6			ESP	BARE
10	560	1	MW BIT	3.0			ESP	SPIRAL
11	550	2	SUB C	0.4		OFA	ESP	CFS
12	550	2	MW BIT	0.7-2.8	YES	OFA	ESP	SPIRAL

[a] Coal Type
MW BIT - Midwestern Bituminous
SUB A, B, C - Subbituminous A, B, C

[b] Gas Cleanup
ESP - Electrostatic Precipitator
FF - Fabric Filter

[c] Economizer
SPIRAL - Spiral Finned
CFS - Continuous Finned
BARE - Bare Tube

NA - Not Available

TABLE 13. CANDIDATE RETROFIT UNITS: 350-475 MWe

			Coal[a]		Gas Cleanup[b]			Type of
Ref. #	Size, MW	No. of Units	Type	% Sulf	FGD	NO_x	Partic	Econ[c]
13	400	1	MW BIT	3.9			ESP	SPIRAL
14	475	1	MW BIT	3.4			ESP	CFS
15	450	1	MW BIT	3.2			ESP	CFS
16	450	1	MW BIT	3.7			NA	CFS
17	450	1	MW BIT	3.2			NA	SPIRAL
18	350	1	SUB C	0.5			ESP	BARE
19	400	2	MW BIT	0.6	YES(1) NA(1)	OFA	ESP	SPIRAL
20	450	2	MW BIT	3.5	YES	OFA	ESP	SPIRAL
21	400	2	MW BIT	0.55	YES	OFA	ESP	SPIRAL
22	450	2	MW BIT	NA	YES	OFA	ESP	SPIRAL
23	350	1	MW BIT	0.6	YES	OFA	ESP	SPIRAL
24	450	1	MW BIT	3.5	YES	OFA	ESP	CFS
25	350	2	SUB B	NA	YES	OFA	FF	BARE

[a] Coal Type
MW BIT - Midwestern Bituminous
SUB A, B, C - Subbituminous A, B, C

[b] Gas Cleanup
ESP - Electrostatic Precipitator
FF - Fabric Filter

[c] Economizer
SPIRAL - Spiral Finned
CFS - Continuous Finned
BARE - Bare Tube

NA - Not Available

TABLE 14. CANDIDATE RETROFIT UNITS: 200-265 MWe

			Coal[a]		Gas Cleanup[b]			
Ref. #	Size, MW	No. of Units	Type	% Sulf	FGD	NO_x	Partic	Type of Econ[c]
26	200	1	SUB B	0.5			ESP	CFS
27	200	1	MW BIT	3.2			ESP	CFS
28	200	1	MW BIT	NA			ESP	CFS
29	265	1	MW BIT	3-4.5			NA	CFS
30	250	1	SUB A	0.5		OFA	ESP	SPIRAL
31	250	2	MW BIT	0.6	YES	OFA	ESP	SPIRAL
32	200	1	MW BIT	2.8	YES	OFA	ESP	SPIRAL
33	255	1	SUB C	NA	YES	OFA	FF	NA
34	265	1	SUB C	NA	YES	OFA	FF	BARE

[a] Coal Type
MW BIT - Midwestern Bituminous
SUB A, B, C - Subbituminous A, B, C

[b] Gas Cleanup
ESP - Electrostatic Precipitator
FF - Fabric Filter

[c] Economizer
SPIRAL - Spiral Finned
CFS - Continuous Finned
BARE - Bare Tube

NA - Not Available

SECTION 10

GENERIC PROCESS DESIGN

The objective of this task was to define the system and operational modifications required to incorporate sorbent injection capabilities in new and existing (retrofit) steam generators. A total of nine cases (three retrofit and six new units) were considered. Sorbent preparation and flue gas cleanup requirements were defined in addition to required boiler modifications for sorbent injection. Table 15 summarizes these nine cases, plus nine alternative configuration cases for which economic analyses were prepared for comparison purposes.

Selection of the retrofit unit was discussed in Section 9. The operating conditions of the three units are summarized in Table 16. Table 17 presents as-received coal analyses for these units.

The new units were selected to reflect the most recent boiler design practice. As-received analyses of the high-sulfur (3.7%S) and low-sulfur (0.55%S) Midwestern bituminous coals used for these cases are shown in Table 18.

Material balances for the three retrofit cases and six new unit cases are shown in Tables 19 through 27.

10.1 Limestone Preparation

The limestone preparation and delivery system was designed by C-E Raymond to produce the required quantities of 10 to 20 micron limestone for direct feed delivery via dilute phase transport. To meet this nominal particle size requirement, roller mills were specified for all cases. Multiple mills are required for all but one of the cases, new and retrofit. The system is shown schematically in Figure 76. The equipment layout is shown in Figure 77 for

TABLE 15. NEW AND RETROFIT UNITS STUDIED

	Size, MWe	Coal Type (High/Low S)	LIMB %SO_2 Removed	Total %SO_2 Removed	Primary FGD Device	Alternative Configuration
Retrofit	560	High	50	50	None	LNCFS Burners
	450	High	50	50	None	LNCFS Burners
	200	High	50	50	None	LNCFS Burners
New	600	High	50	90	Spray Dryer Scrubber	
	400	High	50	90	Spray Dryer Scrubber	
	200	High	50	90	Spray Dryer Scrubber	
New	600	Low	50	70	Spray Dryer Scrubber	
	400	Low	50	70	Spray Dryer Scrubber	
	200	Low	50	70	Spray Dryer Scrubber	

TABLE 16. OPERATING CONDITIONS FOR RETROFIT UNITS

Size, MWe	560	450	200
Primary steam, pph	3,800,000	3,145,200	1,415,000
Superheat outlet temp., °F	1005	1005	1005
Superheat outlet pressure, psia	2620	2621	2100
Reheat steam, pph	3,330,000	2,767,470	1,258,300
Reheat outlet temp., °F	1005	1000	1005

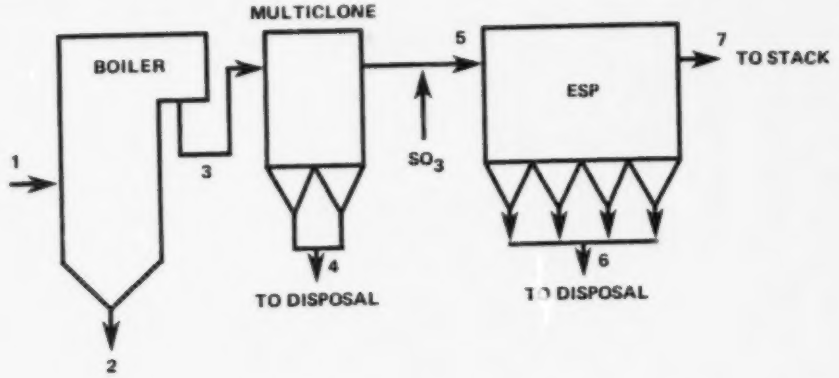
TABLE 17. AS-RECEIVED COAL ANALYSES FOR RETROFIT UNITS

Unit	560 MWe Retrofit	450 MWe Retrofit	200 MWe Retrofit
Proximate analysis, wt %			
Moisture	10.8	9.7	11.8
Volatile matter	33.9	42.3	37.8
Fixed carbon	40.4	39.6	39.8
Ash	14.9	8.4	10.6
Ultimate analysis, wt %			
Moisture	10.8	9.7	11.80
Carbon	62.0	64.4	56.56
Hydrogen	4.4	4.9	5.55
Nitrogen	1.3	1.5	1.76
Sulfur	3.0	3.2	3.22
Oxygen	3.6	7.9	10.51
Ash	14.9	8.4	10.60
HHV, Btu/lb (as-received)	11,150	11,860	11,060
Ash analysis, wt %			
SiO_2 (silica)	37.6	40.1	49.7
Fe_2O_3 (ferric oxide)	28.6	34.4	17.7
Al_2O_3 (alumina)	17.0	21.0	20.4
TiO_2 (titanium oxide)	1.0	N/A	0.8
CaO (calcium oxide)	4.8	1.0	3.4
MgO (magnesia)	0.7	N/A	1.2
SO_3 (sulfur trioxide)	8.5	N/A	3.8
K_2O (potassium oxide)	1.3	N/A	0.1
Na_2O (sodium oxide)	0.1	N/A	N/A
P_2O_5 (phosphorous pentoxide)	N/A	N/A	0.1
Undetermined	0.4	3.5	2.8
Total	100.0	100.0	100.0

TABLE 18. AS-RECEIVED COAL ANALYSES FOR NEW UNITS

Unit	New Units High-Sulfur	New Units Low-Sulfur
Proximate analysis, wt %		
Moisture	16.50	6.43
Volatile matter	34.74	40.67
Fixed carbon	39.36	44.08
Ash	9.40	8.82
Ultimate analysis, wt %		
Moisture	16.50	6.43
Carbon	57.72	68.30
Hydrogen	4.07	4.91
Nitrogen	1.03	0.81
Sulfur	3.70	0.55
Oxygen	7.58	10.18
Ash	9.40	8.82
HHV, Btu/lb (as-received)	10,550	12,192
Ash analysis, wt %		
SiO_2 (silica)	49.60	52.70
Fe_2O_3 (ferric oxide)	19.00	9.88
Al_2O_3 (alumina)	18.30	17.63
TiO_2 (titanium oxide)	0.90	1.45
CaO (calcium oxide)	4.40	9.69
MgO (magnesia)	1.00	1.84
SO_3 (sulfur trioxide)	2.90	0.71
K_2O (potassium oxide)	2.10	0.66
Na_2O (sodium oxide)	1.30	1.33
P_2O_5 (phosphorous pentoxide)	0.40	N/A
Undetermined	0.10	4.11
Total	100.00	100.00

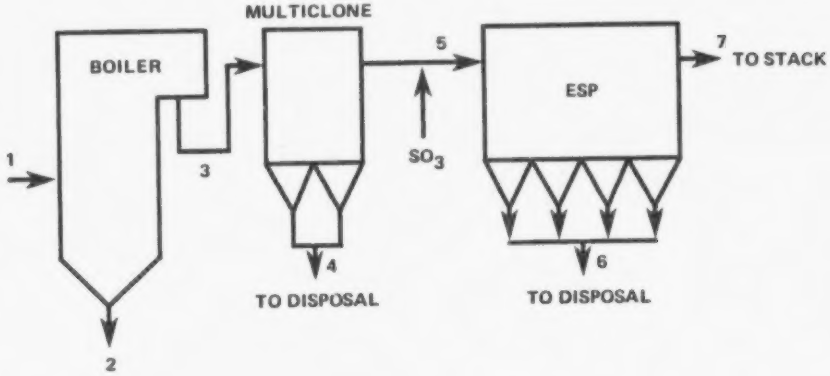
TABLE 19. MATERIAL BALANCE: 560 MWe RETROFIT UNIT


MULTICLONE
BOILER
1
2
3
4
TO DISPOSAL
5
SO_3
ESP
6
TO DISPOSAL
7
TO STACK

Stream (lb/hr) \ Station	1	2*	3	4	5	6	7**
Coal	448,843						
Limestone	88,481						
Air	4,706,110						
CaO			35,326	19,783	15,543	15,398	145
$CaSO_4$			28,588	16,009	12,579	12,462	117
Ash & inerts		13,376	57,926	32,439	25,487	25,249	238
Flue gas			5,125,000				
SO_3					25.6		

* Ash only
** Particulate emission limit: 0.1 lb/10^6 Btu

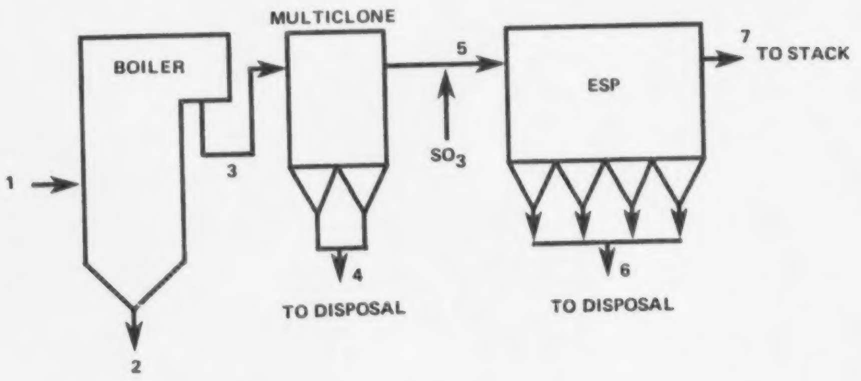
TABLE 20. MATERIAL BALANCE: 450 MWe RETROFIT UNIT


MULTICLONE
BOILER
ESP
1
2
3
4
5
6
7
SO_3
TO STACK
TO DISPOSAL
TO DISPOSAL

Stream (lb/hr) \ Station	1	2*	3	4	5	6	7**
Coal	349,880						
Limestone	73,571						
Air	3,783,720						
CaO			29,373	20,796	8,527	8,425	152
$CaSO_4$			23,769	16,828	6,941	6,818	123
Ash & inerts		5,878	27,191	19,251	7,940	7,800	140
Flue gas			4,100,000				
SO_3					20.5		

* Ash only
** Particulate emission limit: 0.1 lb/10^6 Btu

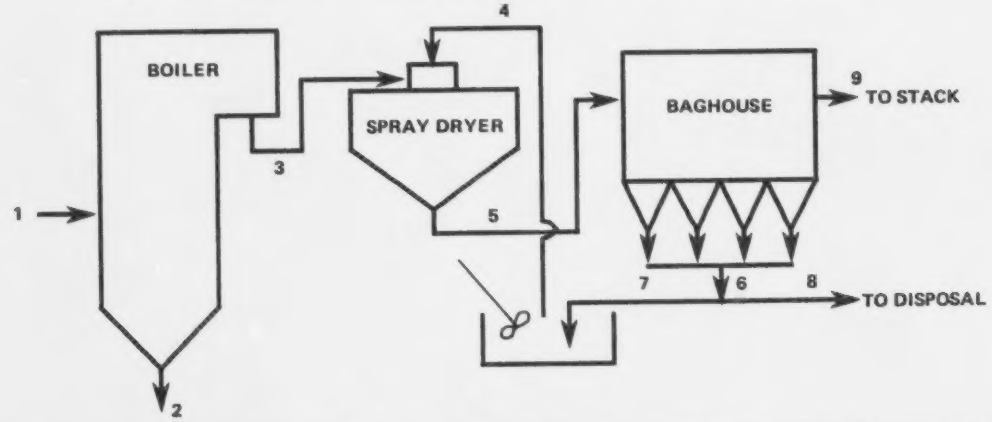
TABLE 21. MATERIAL BALANCE: 200 MWe RETROFIT UNIT


BOILER
MULTICLONE
ESP
1
2
3
4
5
6
7
SO_3
TO DISPOSAL
TO DISPOSAL
TO STACK

Stream (lb/hr) \ Station	1	2*	3	4	5	6	7**
Coal	196,044						
Limestone	41,481						
Air	1,931,226						
CaO			16,561	10,914	5,647	5,573	74
$CaSO_4$			13,402	8,832	4,570	4,510	60
Ash & inerts		4,156	18,698	12,321	6,377	6,294	83
Flue gas			2,130,000				
SO_3					10.7		

* Ash only
** Particulate emission limit: 0.1 lb/10^6 Btu

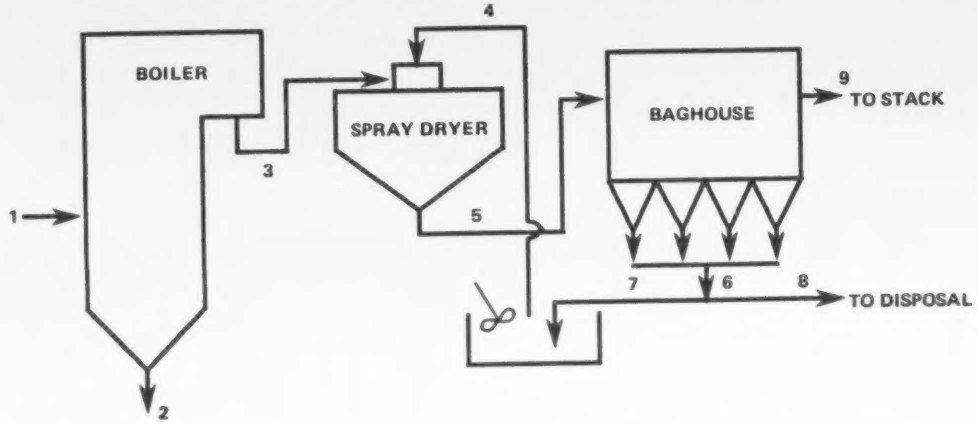
TABLE 22. MATERIAL BALANCE: 600 MWe HIGH-SULFUR NEW UNIT


BOILER
SPRAY DRYER
BAGHOUSE
1
2
3
4
5
6
7
8
9
TO STACK
TO DISPOSAL

Stream (lb/hr) \ Station	1	2*	3	4	5	6	7	8	9**
Coal	523,084								
Limestone	127,177								
Air	5,006,968								
CaO			50,776		50,776	50,744	20,585	30,159	32
$CaSO_4$			41,089	30,257	74,633	74,586	30,257	44,329	47
$CaSO_3$				17,800	43,907	43,879	17,800	26,079	28
$Ca(OH)_2$				33,540	15,644	15,634	6,342	9,292	10
Ash & inerts		9,834	45,695	31,156	76,851	76,802	31,156	45,646	49
Water				220,787					
Flue gas			5,490,692						

* Ash only
** Particulate emission limit: 0.03 lb/10^6 Btu

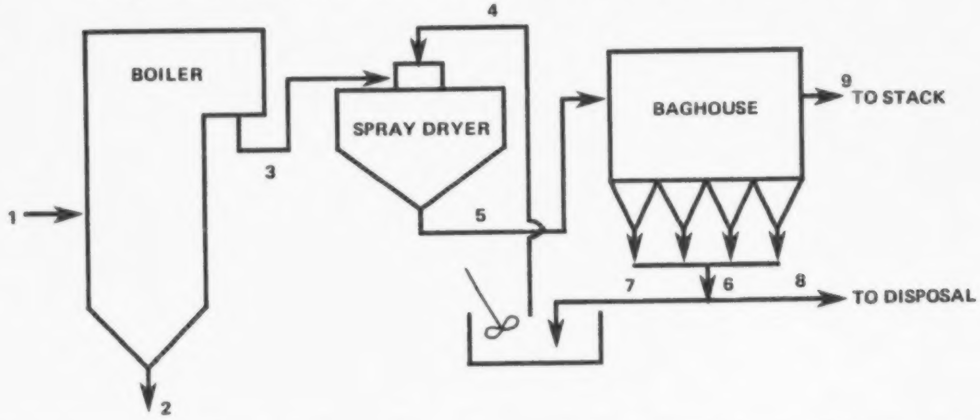
TABLE 23. MATERIAL BALANCE: 400 MWe HIGH-SULFUR NEW UNIT


BOILER
SPRAY DRYER
BAGHOUSE
1
2
3
4
5
6
7
8
9
TO STACK
TO DISPOSAL

Stream (lb/hr) / Station	1	2*	3	4	5	6	7	8	9**
Coal	348,722								
Limestone	84,785								
Air	3,337,978								
CaO			33,851		33,851	33,829	13,723	20,107	21
$CaSO_4$			27,393	20,171	49,755	49,724	20,171	29,553	31
$CaSO_3$				11,867	29,271	29,253	11,867	17,385	19
$Ca(OH)_2$				22,360	10,429	10,422	4,228	6,194	7
Ash & inerts		6,556	30,463	20,771	51,234	51,201	20,770	30,431	33
Water				147,191					
Flue gas			3,660,461						

* Ash only
** Particulate emission limit: 0.03 lb/10^6 Btu

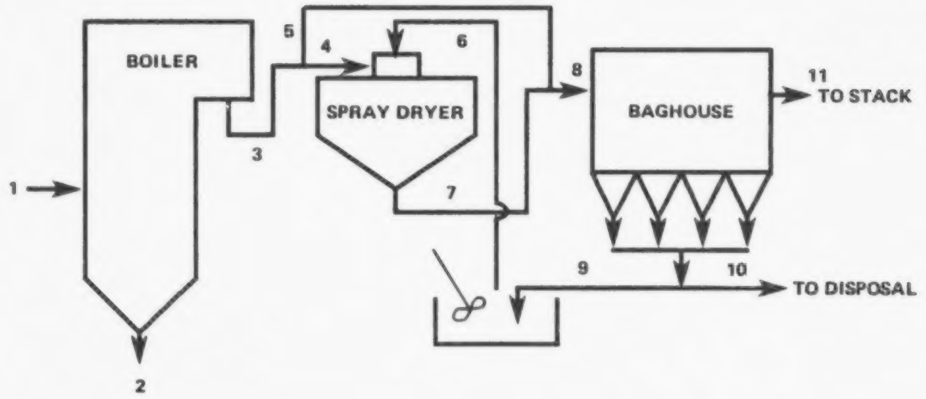
TABLE 24. MATERIAL BALANCE: 200 MWe HIGH-SULFUR NEW UNIT


BOILER
SPRAY DRYER
BAGHOUSE
1
2
3
4
5
6
7
8
9
TO STACK
TO DISPOSAL

Stream (lb/hr) \ Station	1	2*	3	4	5	6	7	8	9**
Coal	174,361								
Limestone	42,392								
Air	1,668,989								
CaO			16,925		16,925	16,914	6,862	10,052	11
$CaSO_4$			13,696	10,086	24,878	24,862	10,086	14,776	16
$CaSO_3$				5,933	14,636	14,627	5,933	8,694	9
$Ca(OH)_2$				11,180	5,215	5,212	2,114	3,098	3
Ash & inerts		3,278	15,232	10,385	25,617	25,601	10,385	15,216	16
Water				73,596					
Flue gas			1,830,230						

* Ash only
** Particulate emission limit: 0.03 lb/10^6 Btu

TABLE 25. MATERIAL BALANCE: 600 MWe LOW-SULFUR NEW UNIT

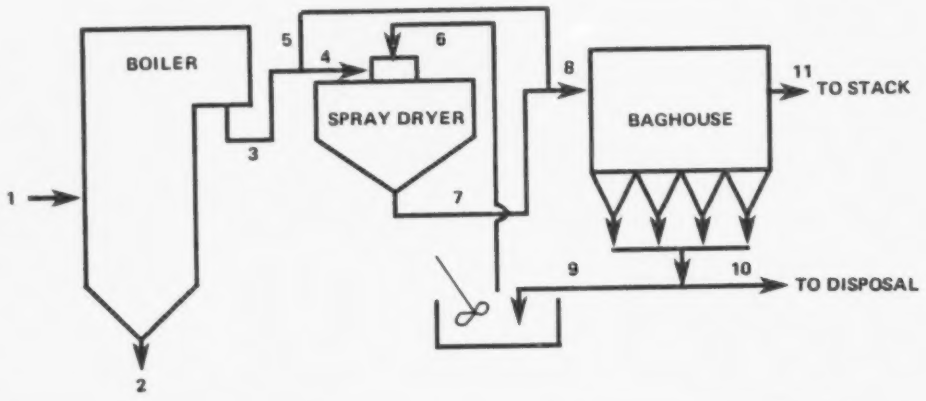

BOILER
SPRAY DRYER
BAGHOUSE
1
2
3
4
5
6
7
8
9
10
11
TO STACK
TO DISPOSAL

Station / Stream (lb/hr)	1	2*	3	4	5	6	7	8	9	10	11**
Coal	451,510										
Limestone	32,636										
Air	4,989,559										
CaO			15,202	7,601	7,601		7,601	15,202	1,551	13,610	41
$CaSO_4$			5,272	2,636	2,636	559	3,406	6,042	559	5,467	16
$CaSO_3$						171	1,846	1,846	171	1,670	5
$Ca(OH)_2$						2,151	1,001	1,001	102	896	3
Ash & inerts		7,965	33,492	16,746	16,746	3,416	20,162	36,908	3,416	33,392	100
Water						119,212					
Flue gas			5,637,817								

* Ash only

** Particulate emission limit: 0.03 lb/10^6 Btu

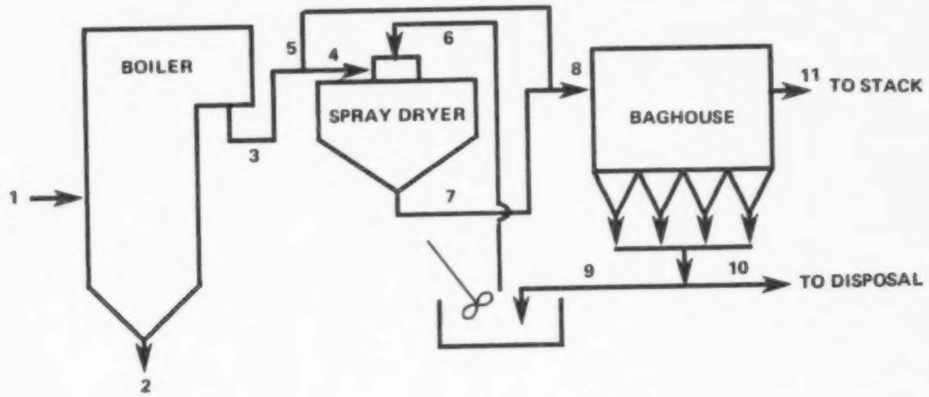
TABLE 26. MATERIAL BALANCE: 400 MWe LOW-SULFUR NEW UNIT


BOILER
SPRAY DRYER
BAGHOUSE
TO STACK
TO DISPOSAL
1
2
3
4
5
6
7
8
9
10
11

Station / Stream (lb/hr)	1	2*	3	4	5	6	7	8	9	10	11**
Coal	301,006										
Limestone	21,757										
Air	3,326,372										
CaO			10,135	5,067	5,068		5,067	10,135	1,034	9,073	27
$CaSO_4$			3,515	1,757	1,758	373	2,271	4,028	373	3,645	11
$CaSO_3$						114	1,231	1,231	114	1,113	3
$Ca(OH)_2$						1,434	667	667	68	597	2
Ash & inerts		5,310	22,328	11,164	11,164	2,277	13,441	24,605	2,277	22,261	67
Water						79,475					
Flue gas			3,758,544								

* Ash only
** Particulate emission limit: 0.03 lb/10^6 Btu

TABLE 27. MATERIAL BALANCE: 200 MWe LOW-SULFUR NEW UNIT


BOILER
SPRAY DRYER
BAGHOUSE
1
2
3
4
5
6
7
8
9
10
11
TO STACK
TO DISPOSAL

Stream (lb/hr) \ Station	1	2*	3	4	5	6	7	8	9	10	11**
Coal	150,503										
Limestone	10,879										
Air	1,663,186										
CaO			5,067	2,534	2,533		2,534	5,067	517	4,537	14
$CaSO_4$			1,757	879	878	186	1,135	2,014	186	1,822	5
$CaSO_3$						57	615	615	57	557	2
$Ca(OH)_2$						717	334	334	34	299	1
Ash & inerts		2,655	11,164	5,582	5,582	1,139	6,721	12,303	1,139	11,131	33
Water						39,737					
Flue gas			1,879,272								

* Ash only
** Particulate emission limit: 0.03 lb/10^6

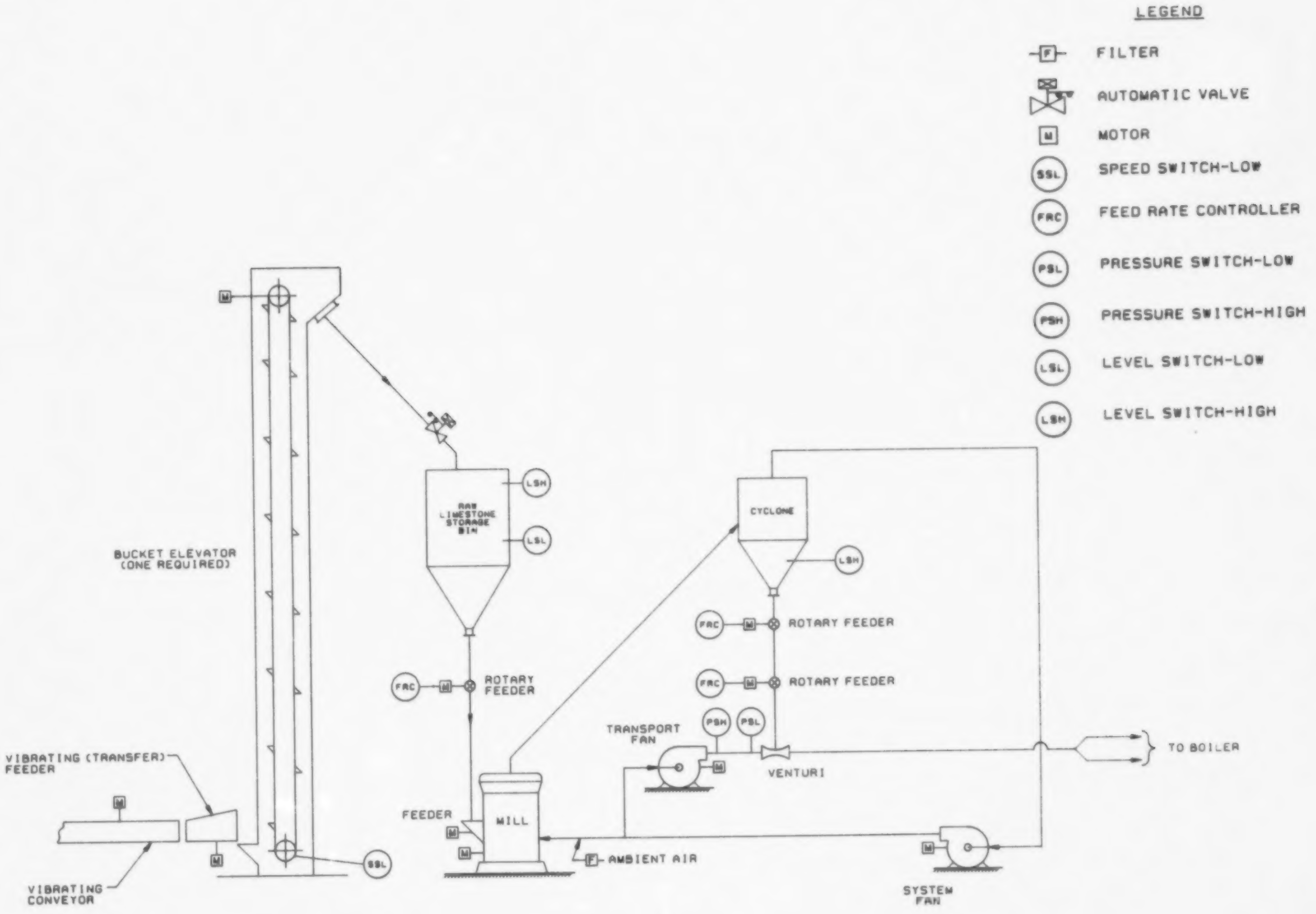
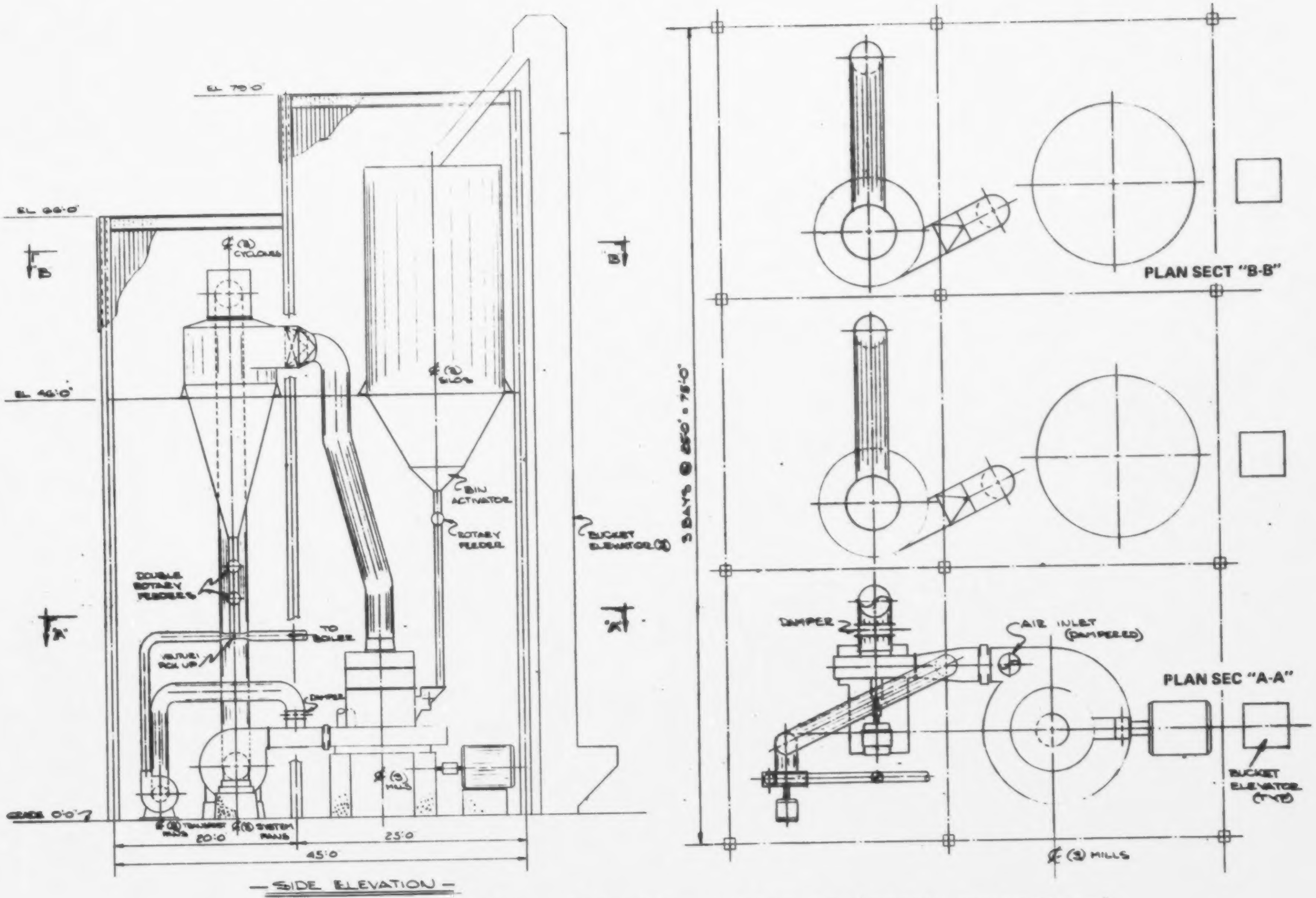

LEGEND
FILTER
AUTOMATIC VALVE
MOTOR
SPEED SWITCH-LOW
FEED RATE CONTROLLER
PRESSURE SWITCH-LOW
PRESSURE SWITCH-HIGH
LEVEL SWITCH-LOW
LEVEL SWITCH-HIGH
F
M
SSL
FRC
PSL
PSH
LSL
LSH
BUCKET ELEVATOR (ONE REQUIRED)
VIBRATING (TRANSFER) FEEDER
VIBRATING CONVEYOR
RAW LIMESTONE STORAGE BIN
ROTARY FEEDER
FEEDER
MILL
AMBIENT AIR
CYCLONE
TRANSPORT FAN
VENTURI
SYSTEM FAN
TO BOILER

Figure 76. Flow schematic for limestone injection system.



— SIDE ELEVATION —

PLAN SECT "B-B"

PLAN SEC "A-A"

EL 70'0"
EL 66'0"
EL 46'0"
GRADE 0'0"
℄ (3) CYCLONES
℄ (3) SILOS
BIN ACTIVATOR
ROTARY FEEDER
BUCKET ELEVATOR (3)
DOUBLE ROTARY FEEDERS
TO BOILER
VENTURI PICK UP
DAMPER
℄ (3) MILLS
℄ (3) TRANSPORT FANS
℄ (3) SYSTEM FANS
20'0"
25'0"
45'0"
B
A
3 BAYS @ 25'0" = 75'0"
DAMPER
AIR INLET (DAMPERED)
BUCKET ELEVATOR (TYP)
℄ (3) MILLS

Figure 77. Limestone injection system equipment arrangement (450 MWe retrofit unit).

the 450 MWe retrofit unit. This section arrangement is based on similar direct feed systems designed by Raymond for applications with pulverized coal and other materials. This arrangement applies to all nine cases, new and retrofit, with the variations being in the number of mill systems and the equipment sizes. Table 28 presents the number and sizes of mills specified for the new and retrofit cases.

The raw limestone is delivered to a storage bin (1 per mill) via a bucket elevator fed by a conveyor and a vibrating feeder. A raw limestone nominal size of 1/2" was assumed. The bin is sized for 12 hours capacity. From each bin, the limestone is fed via a variable-speed rotary feeder to the mill which is capable of producing 99+% through 325 mesh fineness. This material is transported to a high efficiency cyclone which removes in excess of 98.5% of the material. The remaining fines are carried through the system fan (exhauster) back to the mill. Just upstream of the mill, at a point of positive pressure, the required quantity of transport air (1.5 lb/lb limestone) is drawn off by the transport fan. This air is replaced downstream at a point of slightly negative pressure by outside ambient air. Flow through both the transport air and replacement air lines is regulated by dampers. The transport fan outlet is at a nominal 25" w.g. pressure. This stream entrains the limestone in a venturi which is fed by a double rotary feeder system. This feeder arrangement protects against the entire feed line being positively pressurized. From the venturi, the material is delivered to the boiler where it splits to feed the injection nozzles. It is presently felt that each mill feed line should split twice which defines the number of nozzle penetrations for each unit (e.g., two mills feed eight nozzles).

10.2 Boiler Modifications: Retrofit Units

Each of the three retrofit units was analyzed to define the modifications which would be required to incorporate sorbent injection for a 50% SO_2 removal. Side elevation drawings of the three units are presented in Figures 78 through 80.

TABLE 28. LIMESTONE PULVERIZER REQUIREMENTS[a]

Unit		Coal Type	Ca/S Ratio	Limestone Flow, lb/hr	(Number)/Size[b] of Mills
Retrofit	560 MWe	High Sulfur	2:1	88,481	(2) 73", (2) 66"
	450 MWe	High Sulfur	2:1	73,571	(2) 73", (1) 66"
	200 MWe	High Sulfur	2:1	41,481	(1) 73", (1) 66"
New	600 MWe	High Sulfur	2:1	127,177	(5) 73"
	400 MWe	High Sulfur	2:1	84,785	(2) 73", (2) 66"
	200 MWe	High Sulfur	2:1	42,392	(1) 73", (1) 66"
New	600 MWe	Low Sulfur	4:1	32,636	(2) 66"
	400 MWe	Low Sulfur	4:1	21,757	(2) 54"
	200 MWe	Low Sulfur	4:1	10,879	(1) 54"

[a] For design purposes and economic analyses, limestone was assumed to be 95% $CaCO_3$ and 5% inerts (impurities).

[b] Mill sizes correspond to the following C-E Raymond Roller Mill model numbers:

73" #73612
66" #6669
54" #5448

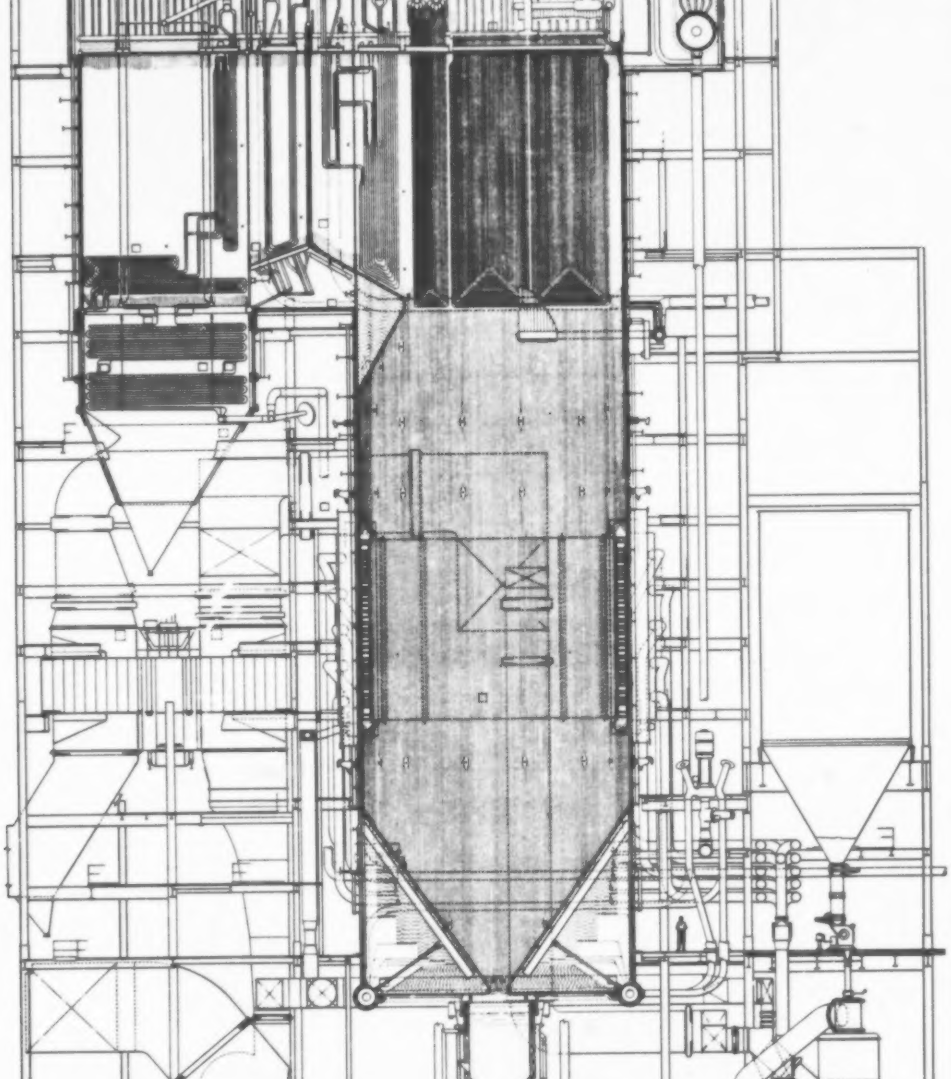


Figure 78. Side elevation drawing, 560 MWe retrofit unit.

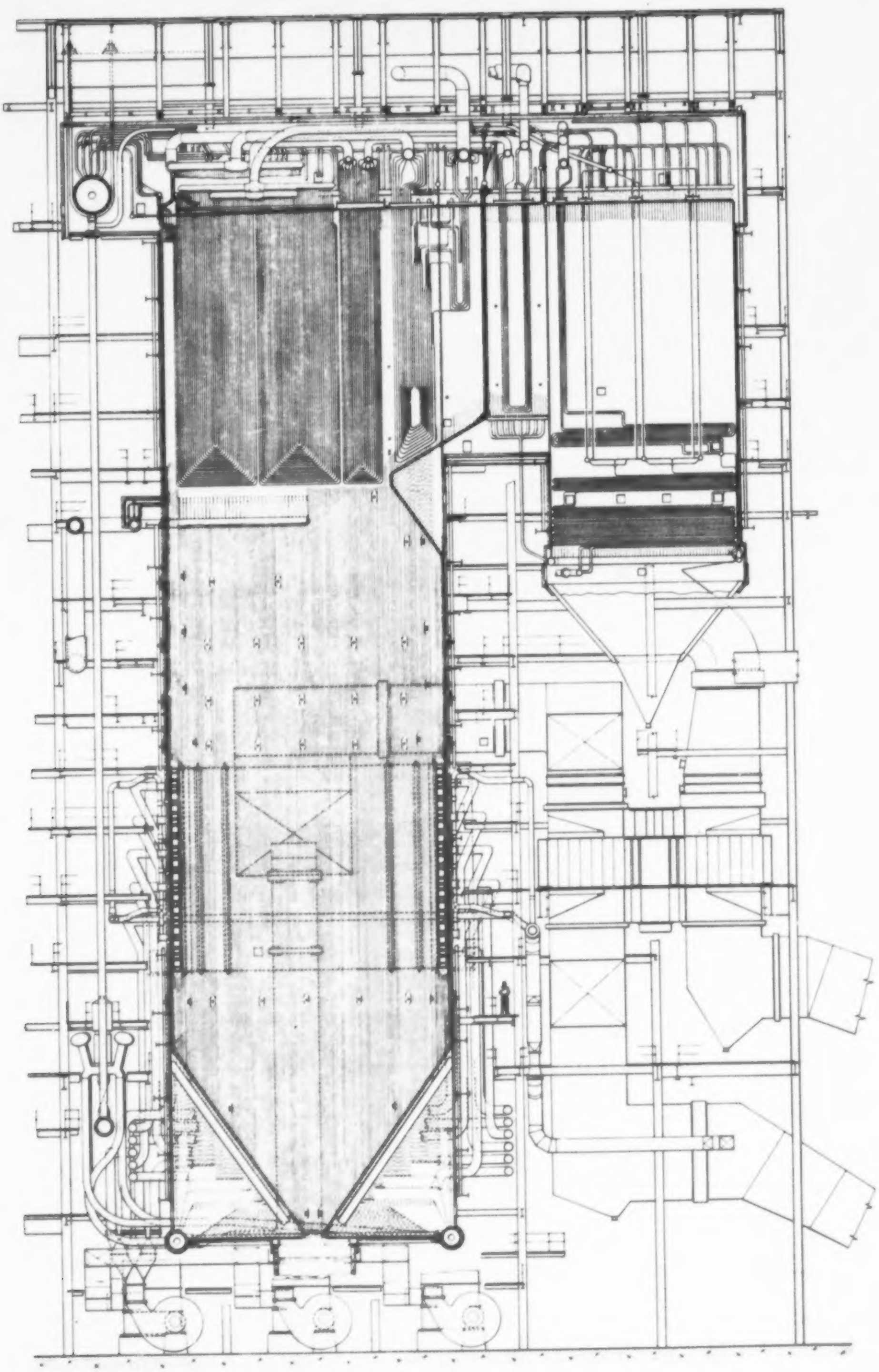


Figure 79. Side elevation drawing, 450 MWe retrofit unit.

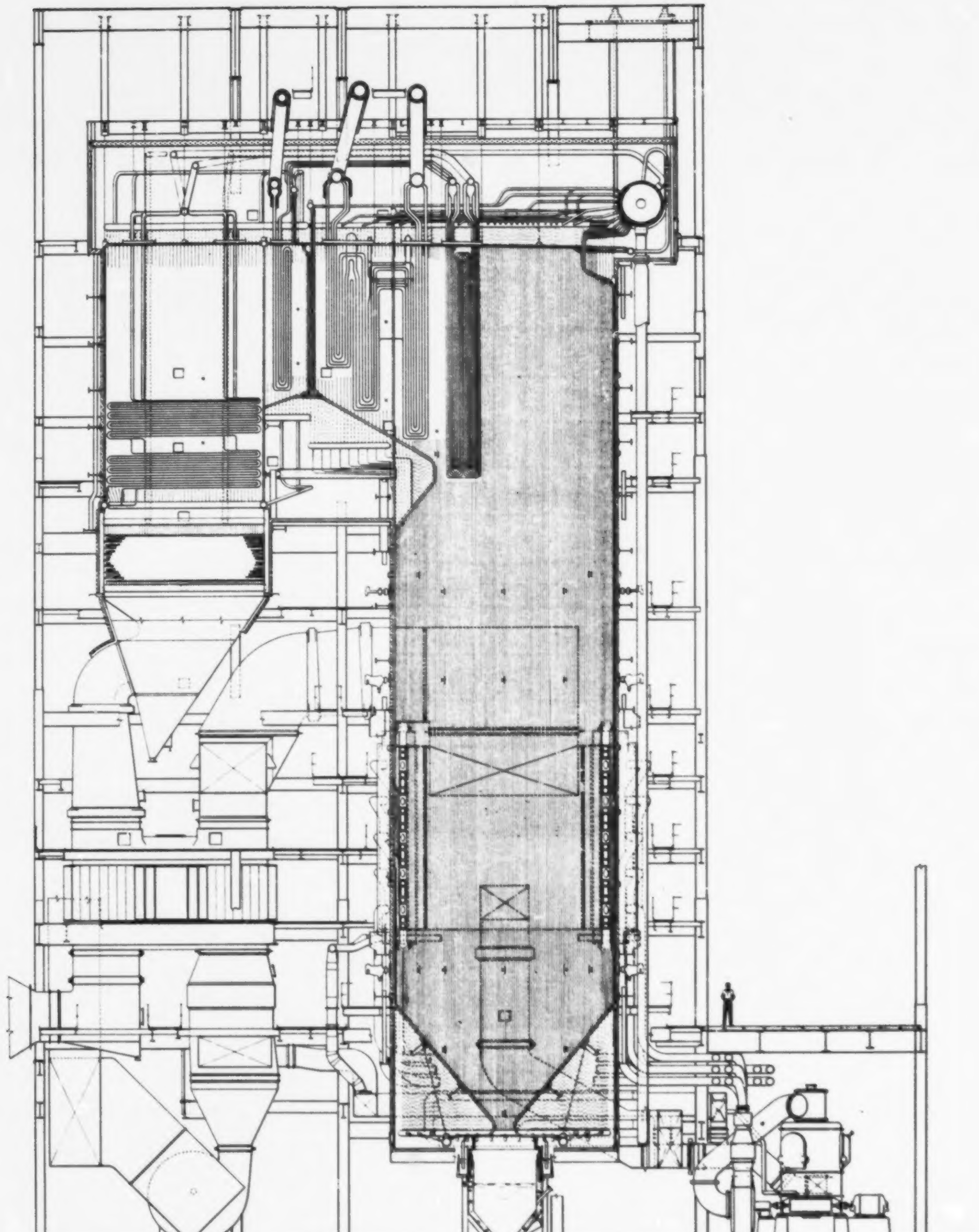


Figure 80. Side elevation drawing, 200 MWe retrofit unit.

The extent of modifications to the candidate units was limited to three areas:

- Replacement of continuous finned/staggered economizers with spiral finned/in-line economizer;
- Addition of sootblowers in economizer and furnace areas;
- Creation of wall penetrations for sorbent injection nozzles.

All three units are presently equipped with economizers with continuous finned tubes in a staggered arrangement. Analyses were made to define the requirements for incorporating the optimum bare tube economizer in an in-line arrangement. Insufficient space was available on any of the retrofit units for this modification. To accomplish the required absorption, and still reconfigure to an in-line arrangement to reduce the potential for plugging, spiral finned tubes were specified, to be incorporated in conjunction with an increase in the number of sootblowers and cleaning frequency.

Nine additional double-sided retractable sootblowers will be required for the 560 MWe unit (7 in the back pass, 2 in the convective pass), 7 will be required for the 450 MWe unit (5 in the back pass, 1 in the convective pass, and one near the bottom of the panels), and 6 will be required for the 200 MWe unit (all in the back pass). This increased cleaning capacity, together with an increase in the frequency of sootblowing, should preclude significant plugging problems during normal operation. It would be recommended that a unit so retrofitted be cleaned on shutdown to remove any accumulated sorbent-based material.

It is anticipated that gas velocities through some convective sections may approach the generally acceptable upper limit of 65 to 70 fps on these units. This is not substantially different with or without sorbent injection. While these velocities may be acceptable under normal firing conditions, some additional erosion might be expected due to the entrained sorbent material. This potential erosion problem was not considered to be severe enough to warrant derating of the units.

The temperature profiles for all three units were such as to suggest a point of injection for the sorbent material at an elevation near the bottom of the panels in the upper furnace area. The maximum temperature at this point is below 2400°F. The sorbent will be injected into this region through multiple nozzles in the walls at a discharge velocity of 200 fps. The average velocity of the gas into which the sorbent is injected is about 45 fps. The behavior of the sorbent material after it is introduced in this manner is subject to question. As discussed in Section 4, tests by EER on the large water-tube simulator furnace on Indiana I coal (2.57%S) showed 37% SO_2 capture at 80 fps injection velocity and 47% SO_2 capture at 100 fps at a Ca/S molar ratio of 2:1 when the sorbent was injected through tertiary air ports. Theoretical analyses would suggest only a short penetration of a few inches at the wall for 20 micron particles injected at 100 fps. The natural convection patterns of the gas in this region would carry the sorbent downward with the cooler wall gas until it is eventually caught up by the hotter gas toward the upper furnace and convective pass areas. Because of this mixing uncertainty, it is difficult to accurately define the overall residence time in the critical 2200°F to 1400°F region, but it is estimated to be between 1.2 and 1.5 seconds. The SO_2 absorption was set at 50% for this study, with 100% calcination and no recarbonation. This defined the quantity and composition of the solids passing through the boiler.

Creating the opening for the sorbent injection nozzles will require moving some waterwall tubes. The average exit diameter of each sorbent injection nozzle is about 6.5". This will require approximately a 12" opening, which, in turn, requires moving approximately 6 waterwall tubes.

No modifications to the ash collection (hopper) and removal equipment on the boiler will be required, since it was defined for purposes of the study that all sorbent-related solids will pass through the boiler [illegible] the particulate removal device. The coal ash split was defined to be 20% boiler/80% ESP with and without sorbent injection.

10.3 Boiler Modifications: New Units

The new units selected for analysis represent current boiler design philosophy with respect to furnace and convective pass sizing. Recent practice reflects a somewhat more conservative approach than that which was employed just 20 years ago. This has resulted in larger furnaces and wider convective passes.

With the 50% SO_2 absorption in the boiler assumed for this study, the boiler modifications required to incorporate limestone injection on new units were relatively minor. The limestone injection point was assumed to be near the bottom of the panels for both high- and low- sulfur units. No modifications to the furnace/convective pass shell were required. Some surface rearranging was recommended to achieve the desired 1.5 second residence time between 2200°F and 1400°F. As discussed in the previous subsection, the behavior of the sorbent material after injection through wall nozzles is uncertain, thus making an accurate determination of solids residence time impossible.

The high-sulfur coal units required the least extensive modifications. This was due in part to the assumption that a substantial reduction in the air heater outlet temperature could be realized due to a decrease in the concentration of SO_3 in the exit gas stream. Consequently, this temperature was reduced to 255°F from 280°F. This change required a change to a larger Ljungstrom air heater which had a significant effect on the overall cost of the modifications. Only small changes in heat transfer surface were required in the finishing superheater, the low temperature superheater, and the economizer. Additional sootblowers were specified for the back pass area to handle the anticipated additional accumulation. Since current practice for economizer design is to specify bare tubes, the impact of reconfiguring to an inline arrangement will not be significant. Velocities through the convective sections did not exceed 50 fps, with or without limestone. While this is substantially below the acceptable upper limits of 65 to 70 fps, additional erosion is expected with limestone injection. Since the severity of this anticipated problem is unknown, no modifications were made to reduce velocities.

Due to the fact that the air heater exit temperature was already at a relatively low 255°F for the base condition, the low-sulfur cases required more extensive modifications to the base design to accommodate limestone injection. Changes were made in the finishing reheat, finishing superheat, low temperature superheat and economizer sections. Because of the relatively small reduction in the amount of preheated secondary air which was replaced by limestone transport air, plus the lack of change in the gas exit temperatures, no change was required to the Ljungstrom air heaters. Additional sootblowers were incorporated, primarily in the rear pass (low temperature) areas. As with the high-sulfur units, the base configuration for the low-sulfur unit incorporates a bare tube economizer. The effect of reconfiguring to an inline arrangement will not be significant. Velocities through some of the low temperature sections approached 60 fps without limestone, still below the acceptable upper limits. To keep velocities at the same levels and still accomplish the same absorptions, tube diameters and arrangement spacings were increased for the finishing and low temperature superheater sections. The finishing reheater tubes were also increased to preclude a substantial increase in pressure drop across this section.

The changes discussed above are summarized in Table 29 for the 600 MWe units. Similar but proportionally smaller changes would be required for 400 MWe and 200 MWe units.

Higher SO_2 removal requirements, different sorbents, and substantially different coal compositions, individually or in combination, could result in substantial changes to the modifications defined in this section.

10.4 Back End Modifications: Retrofit Units

Each of the retrofit units considered for this study is equipped with an electrostatic precipitator (ESP) for particulate control. The 450 MWe unit, which was selected as the candidate host site, is presently equipped with two ESP's. The second higher capacity ESP, which is the only one currently in use, was added when the unit began burning a high-fouling coal. The original

TABLE 29: SUGGESTED BOILER MODIFICATIONS FOR LIMB APPLICATION ON A NEW 600 MWe UNIT

(50% SO_2 CAPTURE)

Unit	Section	Area (ft^2) w/o LIMB	Area (ft^2) w/LIMB	No. Additional Sootblowers	Air Heater
Low-Sulfur (0.55% S)	RH Finish	13,451	17,361	24 Retract.	No Change
	SH Finish	33,047	32,942	4 Wall	
	Low Temp SH Pendant	12,317	11,227		
	Low Temp SH Horizontal	107,484	127,864		
	Economizer	130,284	127,470		
High-Sulfur (3.7% S)	RH Finish	16,006	17,747	18 Retract.	Upgraded to (2)33 VI 74 from (2)31 VI 64
	Low Temp SH Horizontal	118,554	121,969	4 Wall	
	Economizer	117,099	115,953		

ESP is no longer on line and serves only as a "knock-out box" upstream of the present ESP, mechanically removing a relatively small amount of flyash. To maintain consistency in the suggested modifications among the three units, this second ESP was not considered to be available.

Several methods were considered for removing the added sorbent-related particulate matter for retrofit application. THese methods included multiclone plus SO_3 conditioning, ESP upgrade, and addition of a fabric filter downstream of the airheater. The fabric filter option was rejected due to anticipated high capital cost. The ESP upgrade was also rejected due to anticipated high capital cost, plus uncertainty in the relationship between ESP performance loss due to sorbent injection and the extent of required modifications to recover that performance. The multiclone/SO_3 conditioning combination should permit the units to meet or better the level of emissions which existed prior to the retrofit. The mechanical collectors will remove from 56% to 71% of the total entering solids, depending on unit size, thus reducing the loading entering the ESP to the level it would experience without sorbent injection. The resistivity of the remaining sorbent-laden particulate matter will be substantially poorer than that of sorbent-free fly ash. Conditioning of the gas by injecting small amounts (5 to 20 ppm) of SO_3 into the flue gas upstream of the ESP has been shown to improve resistivity levels, thus offsetting the ESP performance penalty associated with sorbent injection. It was assumed for this study that the finer distribution of the particulate matter leaving the multiclone would not significantly affect ESP removal efficiency.

10.5 Back End Modifications: New Units

Two methods of particulate removal were considered for the new units. As a reference, for new units without sorbent injection, flue gas desulfurization and particulate control via wet scrubbing with oxidation was assumed. Costs for this reference equipment were obtained from the 1983 Tennessee Valley Authority study on flue gas desulfurization costs, Ref. 29. For new units with limestone injection, a spray dryer/baghouse combination was incorporated to remove sufficient additional SO_2 and sufficient particulate to meet NSPS standards (90% SO_2 for high-sulfur; 70% SO_2 for low-sulfur coal; and 0.03 $lb/10^6$ Btu particulate).

The spray dryer system utilized was similar to that discussed in Section 7. Recycled material from the baghouse was utilized at a stoichiometry of 1.5:1. No additional lime requirement was assumed and no solid material was removed in the absorber vessel itself. All particulate removal occurred in the baghouse. Although as much as 10% to 15% SO_2 absorption might be expected to occur in the baghouse, credit for this absorption was not taken in sizing and costing the spray dryer systems. Since the baghouse cost is not attributable to the SO_2 removal system (LIMB plus spray dryer), this was considered to be a valid, albeit conservative, approach. Additional capital and operating costs for additional baghouse solids handling capability on LIMB/spray dryer-equipped units were included in the economic analyses.

SECTION 11

ECONOMIC ANALYSIS

Total costs which were developed for the new and retrofit units are presented in this section. Costs for the retrofit units were developed for two configurations: with and without low-NO_x firing system retrofit. Costs for the new units were developed with LIMB and compared with conventional flue gas desulfurization equipment (scrubber). The EPRI format as outlined in the Technical Assessment Guide (TAG) was used. Specific assumptions for individual cost items were arrived at after discussion with the EPA Project Officer.

11.1 Economic Premises

11.1.1 Capital Investment Basis

Total Plant Cost - Total plant cost (TPC) comprises the following components: process capital, general facilities, engineering and home office overhead and fee, project contingency, and process contingency.

Process capital is the total constructed cost of all on-site processing and operating units, including all direct and indirect construction costs. For this study, the major units included in this category are the limestone preparation and injection system, the back end equipment, and any required boiler modifications.

General facilities include roads, office buildings, shops, laboratories, etc. Generally, an amount in the range of 5% to 20% of the process capital is assumed. For this study, 5% of process capital was used.

Engineering and home office overhead and fee are generally assumed to be in the range of 10% to 15% of the process capital cost. For this study, 10% of process capital was used.

Project contingency is a factor which is intended to cover additional equipment or other costs that result from a more detailed design of a definitive project at an actual site. This factor, as presented in the EPRI TAG, is a function of the degree of accuracy and type of design estimate carried out. The cost estimates for this study are of the Class II (Preliminary) variety, which specifies a project contingency range of 15% to 30% of process capital. For this study, a project contingency of 20% of process capital was used.

The process contingency is applied to new technology in an effort to quantify the uncertainty in the design and cost of commercial scale equipment. This factor is a function of the state of development for a given technology. For a technology which has been demonstrated on a full-size module (e.g., 20 to 100 MWe), a process contingency range of 5% to 20% of process capital is suggested in the TAG. LIMB falls in this state of development category. For this study, a process contingency of 10% of process capital was used.

Total Capital Requirement - The total capital requirement includes all capital necessary to complete the entire project. Included are the following: total plant investment (includes allowance for funds during construction and escalation), preproduction (startup) costs, inventory capital, initial chemical and catalyst charge, land, and prepaid royalties.

Total plant investment (TPI) at the date of startup includes escalation of construction costs and allowance for funds used during construction (AFDC) over the construction period. This factor is applied to TPC which represents overnight or instantaneous construction. For this study, TPC = TPI, representing a January, 1986 startup of a plant constructed in December, 1985.

Preproduction costs are intended to cover operator training, equipment checkout, major changes in plant equipment, extra maintenance, and inefficient use of fuel and other materials during plant startup. The preparation costs included the following:

a) One month fixed operating costs (fixed operating costs are operating and maintenance labor, administrative and support labor, and maintenance materials).

b) One month of variable operating costs at full capacity, excluding fuel. For this study, these variable operating costs include limestone, waste disposal, and, where applicable, SO_3 and water.

c) A fuel cost of 25% of full capacity for one month to cover inefficient operation which occurs during the startup period.

d) A charge of 2% of TPI to cover expected changes in and modifications to equipment which will be needed to bring the plant to full capacity.

The value of inventories of fuel, other consumables, and by-products is capitalized and included in the inventory capital account. For this study, the following items were included:

a) A fuel supply representing 60 days of full load operation.

b) A supply of other consumables (limestone, SO_3), excluding water, representing 60 days of full load operation. In-plant power consumption (ranging from 0.22 to 0.92 percent of rated capacity) is included under this category.

For this study, there were no initial chemicals and catalyst charges, land charges, or prepaid royalties charges.

11.1.2 Operating Cost Basis

The operating costs were estimated on a first year basis in December, 1985 dollars. The "first year" costs assume normal operation and do not include extra startup costs, which are computed separately. The costs were also computed on a 30-year levelized basis (new units) or a 15-year levelized basis (retrofit units).

The operating costs are divided into fixed and variable and consumable costs. The fixed costs are essentially independent of capacity factor and are generally expressed in $/kW-yr. The variable and consumable costs, collectively called marginal operating costs, are directly proportional to the amount of power produced and are generally expressed in mills/kWh.

Operating and Maintenance Costs - Operating costs include operating labor maintenance (fixed and variable) and overhead charges.

The number of operating jobs (OJ) required to operate the LIMB associated equipment was estimated to be 1.5. The average labor rate (ALR) was set at $23.37 hr. The operating labor charges (OLC) were calculated as follows:

$$\text{OLC} = \frac{\text{OJ x ALR x 8760 hr/yr}}{\text{Net capacity of plant at full load in KW}}$$

Annual maintenance costs for new technologies are often estimated as a percentage of the installed capital cost of the facilities. The percentage varies widely depending on the nature of the processing conditions and the type of design. A representative range of values is 1.5% to 10%. For this study, a value of 4% of process capital was used.

Maintenance costs are generally expressed as maintenance labor and maintenance materials. Since detailed actual information was not available, a maintenance labor/materials ratio of 40/60 was used for this study.

From utility experience, base loaded plants tend to have a relatively high fixed component of operating costs. To break the operating costs into fixed and variable components, it was assumed that the fixed portion is equal to the expected nominal capacity factor (65% for base load units). The balance of the cost was assumed to be variable.

The only overhead charge included in these studies is a charge for administrative and support labor. This charge is taken as 30% of the sum of operating and maintenance labor.

Consumables Operating Costs - Consumables costs include limestone, water, SO_3, waste disposal, and in-plant power. Table 30 presents the individual in-plant incremental power requirements for the LIMB-equipped units. This information was obtained from the various equipment manufacturers. The power indicated for ash handling reflects the incremental power requirements to handle the added sorbent-related material. The following charges were used:

limestone cost: $17.88/ton
water cost: $0.77/1000 gal
SO_3 cost: $209/ton sulfur
disposal cost: $7.66/ton (dry)
in-plant power cost: 58 mills/kWh

Fuel Costs - Fuel costs were assumed to be $\$2.61/10^6$ Btu.

By-product Credits - No by-product credits were taken.

Levelized Operating Costs - Inflation will tend to increase the operating costs (in current dollars) over the life of the plant. For this study, a long-term escalation rate of 8.5%/yr is assumed in estimating the cost of capital (discussed below) and in estimating the life cycle revenue requirements for other expenses. To represent those varying revenue requirements for fixed and variable costs (including fuel), a single "levelized" value is computed using the "present worth" concept of money. Based on the assumption of an 8.5%/yr rate of inflation and a 12.5%/yr discount rate, the levelization factors for operating and maintenance costs (excluding fuel) are 2.314 for new units (30 year) and 1.718 for retrofit units (15 year).

11.1.3 Cost of Capital

The cost of capital is based on an assumed 8.5%/yr inflation rate and the following assumptions:

TABLE 30. IN-PLANT INCREMENTAL POWER REQUIREMENTS FOR LIMB-EQUIPPED RETROFIT AND NEW UNITS (KW)

Unit	Mills	Transport Fans	Feeders, Conveyors, Elevators	Ash Handling	Spray Dryer	Total KW	% of Rating
Retrofit							
560 MWe	1625.6	234.2	214.8	44.7	-	2119.3	0.38
450 MWe	1297.5	187.2	161.1	44.7	-	1690.3	0.38
200 MWe	812.8	117.1	107.4	44.7	-	1082.0	0.54
New High-Sulfur							
600 MWe	2423.5	350.5	268.5	44.7	2037.1	5124.3	0.85
400 MWe	1625.6	234.2	214.8	44.7	1569.0	3688.3	0.92
200 MWe	812.8	117.1	107.4	44.7	715.4	1797.4	0.90
New Low-Sulfur							
600 MWe	656.2	94.0	107.4	44.7	589.2	1491.5	0.25
400 MWe	231.2	61.2	107.4	44.7	453.8	898.5	0.22
200 MWe	115.6	30.6	53.7	44.7	206.9	451.5	0.23

Debt/equity ratio		50/50
Debt cost		11.0%/yr
Preferred stock ratio		15%
Preferred stock cost		11.5%/yr
Common stock ratio		35%
Common stock cost		15.3%/yr
Weighted cost of capital		12.5%/yr
Federal and state income tax		50%
Property tax and insurance		2%/yr
Investment tax credit		10%
Book life	(fossil plant-new)	30 yr
	(fossil plant-retrofit)	15 yr
Tax life	(fossil plant-new)	15 yr
	(fossil plant-retrofit)	5 yr

The 30-year levelized fixed charge rate (LFCR) calculated from the above assumptions is 15.3%/yr for new units. The 15-year LFCR calculated from these assumptions is 16.1%/yr for retrofit units.

The levelized fixed charges (LFC) are based on the Total Capital Requirements (TCR) and are computed as follows:

$$\text{LFC} = \frac{\text{LFCR x TCR}}{\text{Plant size, kW}} = \$/\text{kW yr}$$

11.1.4 Busbar Power Cost

Levelized busbar costs of power for the design capacity factor (0.65) were calculated using the 30-year (new) or 15-year (retrofit) levelized fixed charges, O&M charges, and fuel charges.

11.2 Economic Analyses

The economic analyses were carried out as outlined in Subsection 11.1. Costs were developed on a \$/kW, \$/ton SO_2 removed, and mills/kW-hr basis, both first year and levelized. For the three retrofit units, costs were developed with and without the LNCFS burner modification for NO_x reduction to provide a measure of how much this option would add to the cost of incorporating limestone injection. For the six new units, costs were developed for LIMB plus spray dryer and for conventional FGD equipment without sorbent injection to provide a comparison of these two alternative SO_2 control options. The results of these analyses are presented in Tables 31 through 48 and Figures 81 through 86.

Table 49 presents a breakdown of the Process Capital Costs into limestone preparation/injection costs, boiler modification costs, LNCFS burner costs, and back end equipment costs. Further breakdown of these costs are presented in Tables 50 through 53. Substantial portions of the limestone preparation costs shown in Table 49 are associated with the economic requirement that this material be finely pulverized on site. These costs would not be incurred if a different sorbent material, such as $Ca(OH)_2$, was selected. Therefore, these process capital costs are further broken down into mill-related costs (Table 50) and non-mill-related costs (Table 51). Costs of pulverizers and motors, and valves, plus associated structural steel and enclosures, were defined as mill-related costs. All remaining costs, including bins, piping, transport fans and motors, feeders, conveyors, elevators, and controls, plus associated structural steel and enclosures, were defined as non-mill-related costs.

Table 52 presents a breakdown of the process capital costs associated with LIMB-related boiler modifications. The costs associated with adding nozzle penetrations are included for retrofit units only, since LIMB-equipped new units would be designed to incorporate these penetrations. The Air Heater Upgrade cost reflects the design change to a lower gas outlet temperature on new high-sulfur units as a result of a drop in the SO_3 concentration with LIMB (Section 10.3).

Table 53 presents a breakdown of the process capital costs associated with back end equipment modifications and/or additions associated with incorporating sorbent injection. For retrofit units, these costs include costs for multiclones plus incremental ash handling capabilities and

TABLE 31: ECONOMIC ANALYSIS OF
560 MWE RETROFIT WITH LIMB[1]

Capital Investment	10^3 $	$/kW	
Process Capital	18702.9	33.40	
General Facilities	935.2	1.67	
Engineering and Home Office Fees	1870.3	3.34	
Project Contingency	3740.6	6.68	
Process Contingency	1870.3	3.34	
Total Plant Cost (Dec. 1985 dollars)	27119.2	48.43	
Total Plant Investment (Jan. 1986 startup)	27119.2	48.43	
Preproduction Costs	3886.3	6.94	
Inventory Capital	1165.2	2.08	
Total Capital Requirement (Jan. 1986 startup)	32170.7	57.45	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 $	$/kW-yr	
Operating Labor	307.1	0.55	
Maintenance Labor	299.3	0.53	
Maintenance Materials	448.9	0.80	
Administrative and Support Labor	181.9	0.32	
Total O&M First Year	1237.1	2.21	
Fixed O&M First Year	804.1	1.44	
Variable O&M First Year	433.0	0.14	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 $	mills/kW-hr	
Limestone	4504.1	1.41	
In-Plant Power (2119.3 KW)	699.9	0.22	
Other Consumables: SO_3	6.1	0.00	
Waste Disposal	1479.4	0.46	
Total Consumables (excluding fuel)	6689.5	2.10	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	33.9	0.01	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	5150.5	9.20	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	13111.0	4.11	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	13111.0	342.00	$/ton

15-Year Levelized O&M Costs	$10^3 \$$		
15-Year Levelized Fixed O&M	1378.2	2.46	$/kW-yr
15-Year Levelized Variable O&M	742.2	0.23	mills/kW-hr
15-Year Levelized Consumables O&M	11465.8	3.60	mills/kW-hr
15-Year Levelized Carrying Charges	5150.5	9.20	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	18799.3	5.90	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	18799.3	490.39	$/ton

Notes: (1) Coal Sulfur Content: 3.0%
Calcium/Sulfur Mole Ratio: 2:1
SO_2 Removed in Boiler: 50%
Overall SO_2 Removal: 50%

TABLE 32: ECONOMIC ANALYSIS OF 450 MWE RETROFIT WITH LIMB[1]

Capital Investment	$10^3 \$$	$/kW	
Process Capital	13717.4	30.48	
General Facilities	685.9	1.52	
Engineering and Home Office Fees	1371.7	3.05	
Project Contingency	2743.4	6.10	
Process Contingency	1371.7	3.05	
Total Plant Cost (Dec. 1985 dollars)	19890.2	44.20	
Total Plant Investment (Jan. 1986 startup)	19890.2	44.20	
Preproduction Costs	3168.4	7.04	
Inventory Capital	968.8	2.15	
Total Capital Requirement (Jan. 1986 startup)	24027.4	53.39	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	$10^3 \$$	$/kW-yr	
Operating Labor	307.1	0.68	
Maintenance Labor	219.5	0.49	
Maintenance Materials	329.2	0.73	
Administrative and Support Labor	158.0	0.35	
Total O&M First Year	1013.8	2.25	
Fixed O&M First Year	658.9	1.46	
Variable O&M First Year	354.8	0.14	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	$10^3 \$$	mills/kW-hr	
Limestone	3745.1	1.46	
In-Plant Power (1690.5 KW)	558.3	0.22	
Other Consumables: SO_3	4.9	0.00	
Waste Disposal	1230.1	0.48	
Total Consumables (excluding fuel)	5538.4	2.16	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	28.1	0.01	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	3846.8	8.55	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	10427.1	4.07	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	10427.1	327.16	$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	1129.4	2.51	$/kW-yr
15-Year Levelized Variable O&M	608.1	0.24	mills/kW-hr
15-Year Levelized Consumables O&M	9492.8	3.70	mills/kW-hr
15-Year Levelized Carrying Charges	3846.8	8.55	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	15128.2	5.90	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	15128.2	474.60	$/ton

Notes: (1) Coal Sulfur Content: 3.2%
Calcium/Sulfur Mole Ratio: 2:1
SO_2 Removed in Boiler: 50%
Overall SO_2 Removal: 50%

TABLE 33: ECONOMIC ANALYSIS OF 200 MWE RETROFIT WITH LIMB[1]

Capital Investment	10^3 \$	\$/kW	
Process Capital	8665.7	43.33	
General Facilities	433.3	2.17	
Engineering and Home Office Fees	866.6	4.33	
Project Contingency	1733.1	8.67	
Process Contingency	866.6	4.33	
Total Plant Cost (Dec. 1985 dollars)	12565.3	62.83	
Total Plant Investment (Jan. 1986 startup)	12565.3	62.83	
Preproduction Costs	1755.1	8.78	
Inventory Capital	545.8	2.73	
Total Capital Requirement (Jan. 1986 startup)	14866.2	74.33	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 \$	\$/kW-yr	
Operating Labor	307.1	1.54	
Maintenance Labor	138.6	0.69	
Maintenance Materials	208.0	1.04	
Administrative and Support Labor	133.7	0.67	
Total O&M First Year	787.4	3.94	
Fixed O&M First Year	511.8	2.56	
Variable O&M First Year	275.6	0.24	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 \$	mills/kW-hr	
Limestone	2111.6	1.85	
In-Plant Power (1690.5 KW)	357.3	0.31	
Other Consumables: SO_3	2.5	0.00	
Waste Disposal	693.9	0.61	
Total Consumables (excluding fuel)	3165.0	2.78	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	14.8	0.01	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	2380.1	11.90	\$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	6347.3	5.57	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	6347.3	353.18	\$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	877.2	4.39	$/kW-yr
15-Year Levelized Variable O&M	472.4	0.41	mills/kW-hr
15-Year Levelized Consumables O&M	5424.8	4.76	mills/kW-hr
15-Year Levelized Carrying Charges	2380.1	11.90	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	9181.4	8.06	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	9181.4	510.87	$/ton

Notes: (1) Coal Sulfur Content:	3.2%
Calcium/Sulfur Mole Ratio:	2:1
SO_2 Removed in Boiler:	50%
Overall SO_2 Removal:	50%

TABLE 34: ECONOMIC ANALYSIS OF
560 MWE RETROFIT WITH LIMB AND LNCFS BURNERS[1]

Capital Investment	10^3 $	$/kW	
Process Capital	19562.9	34.93	
General Facilities	978.1	1.75	
Engineering and Home Office Fees	1956.3	3.49	
Project Contingency	3912.6	6.99	
Process Contingency	1956.3	3.49	
Total Plant Cost (Dec. 1985 dollars)	28366.2	50.65	
Total Plant Investment (Jan. 1986 startup)	28366.2	50.65	
Preproduction Costs	3914.4	6.99	
Inventory Capital	1165.2	2.08	
Total Capital Requirement (Jan. 1986 startup)	33445.8	59.72	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 $	$/kW-yr	
Operating Labor	307.1	0.55	
Maintenance Labor	313.0	0.56	
Maintenance Materials	469.5	0.84	
Administrative and Support Labor	186.0	0.33	
Total O&M First Year	1275.6	2.28	
Fixed O&M First Year	829.2	1.48	
Variable O&M First Year	446.5	0.14	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 $	mills/kW-hr	
Limestone	4504.1	1.41	
In-Plant Power (2119.3 KW)	699.9	0.22	
Other Consumables: SO_3	6.1	0.00	
Waste Disposal	1479.4	0.46	
Total Consumables (excluding fuel)	6689.5	2.10	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	33.9	0.01	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	5354.7	9.56	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	13353.7	4.19	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	13353.7	348.34	$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	1421.2	2.54	$/kW-yr
15-Year Levelized Variable O&M	765.3	0.24	mills/kW-hr
15-Year Levelized Consumables O&M	11465.8	3.60	mills/kW-hr
15-Year Levelized Carrying Charges	5354.7	9.56	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	19068.6	5.98	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	19068.6	497.41	$/ton

Notes: (1) Coal Sulfur Content:	3.0%
Calcium/Sulfur Mole Ratio:	2:1
SO_2 Removed in Boiler:	50%
Overall SO_2 Removal:	50%

TABLE 35: ECONOMIC ANALYSIS OF
450 MWE RETROFIT WITH LIMB AND LNCFS BURNERS[1]

Capital Investment	$10^{3}\ \$$	$/kW	
Process Capital	14490.4	32.20	
General Facilities	724.5	1.61	
Engineering and Home Office Fees	1449.0	3.22	
Project Contingency	2898.1	6.44	
Process Contingency	1449.0	3.22	
Total Plant Cost (Dec. 1985 dollars)	21011.1	46.69	
Total Plant Investment (Jan. 1986 startup)	21011.1	46.69	
Preproduction Costs	3193.7	7.10	
Inventory Capital	968.8	2.15	
Total Capital Requirement (Jan. 1986 startup)	25173.6	55.94	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	$10^{3}\ \$$	$/kW-yr	
Operating Labor	307.1	0.68	
Maintenance Labor	231.8	0.52	
Maintenance Materials	347.8	0.77	
Administrative and Support Labor	161.7	0.36	
Total O&M First Year	1048.4	2.33	
Fixed O&M First Year	681.4	1.51	
Variable O&M First Year	366.9	0.14	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	$10^{3}\ \$$	mills/kW-hr	
Limestone	3745.1	1.46	
In-Plant Power (1690.5 KW)	558.3	0.22	
Other Consumables: SO_3	4.9	0.00	
Waste Disposal	1230.1	0.48	
Total Consumables (excluding fuel)	5538.4	2.16	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	28.1	0.01	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	4030.3	8.96	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	10645.2	4.15	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	10645.2	333.96	$/ton

15-Year Levelized O&M Costs	$10^3\,\$$		
15-Year Levelized Fixed O&M	1167.9	2.60	$/kW-yr
15-Year Levelized Variable O&M	628.9	0.24	mills/kW-hr
15-Year Levelized Consumables O&M	9492.8	3.70	mills/kW-hr
15-Year Levelized Carrying Charges	4030.3	8.96	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	15371.0	6.00	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	15371.0	482.22	$/ton

Notes: (1) Coal Sulfur Content: 3.2%
Calcium/Sulfur Mole Ratio: 2:1
SO_2 Removed in Boiler: 50%
Overall SO_2 Removal: 50%

TABLE 36: ECONOMIC ANALYSIS OF 200 MWE RETROFIT WITH LIMB AND LNCFS BURNERS[1]

Capital Investment	10^3 $	$/kW	
Process Capital	9272.7	46.36	
General Facilities	463.6	2.32	
Engineering and Home Office Fees	927.3	4.64	
Project Contingency	1854.5	9.27	
Process Contingency	927.3	4.64	
Total Plant Cost (Dec. 1985 dollars)	13445.4	67.23	
Total Plant Investment (Jan. 1986 startup)	13445.4	67.23	
Preproduction Costs	1775.0	8.88	
Inventory Capital	545.8	2.73	
Total Capital Requirement (Jan. 1986 startup)	15766.2	78.83	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 $	$/kW-yr	
Operating Labor	307.1	1.54	
Maintenance Labor	148.4	0.74	
Maintenance Materials	222.5	1.11	
Administrative and Support Labor	136.6	0.68	
Total O&M First Year	814.6	4.07	
Fixed O&M First Year	529.5	2.65	
Variable O&M First Year	285.1	0.25	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 $	mills/kW-hr	
Limestone	2111.6	1.85	
In-Plant Power (1082.0 KW)	357.3	0.31	
Other Consumables: SO_3	2.5	0.00	
Waste Disposal	693.9	0.61	
Total Consumables (excluding fuel)	3165.0	2.78	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	14.8	0.01	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	2524.2	12.62	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	6518.6	5.72	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	6518.6	362.70	$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	907.6	4.54	$/kW-yr
15-Year Levelized Variable O&M	488.7	0.43	mills/kW-hr
15-Year Levelized Consumables O&M	5424.8	4.76	mills/kW-hr
15-Year Levelized Carrying Charges	2524.2	12.62	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	9372.2	8.23	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	9372.2	521.50	$/ton

Notes: (1) Coal Sulfur Content: 3.22%
Calcium/Sulfur Mole Ratio: 2:1
SO_2 Removed in Boiler: 50%
Overall SO_2 Removal: 50%

TABLE 37: ECONOMIC ANALYSIS OF
600 MWE NEW UNIT WITH LIMB AND SPRAY DRYER (HIGH-SULFUR)[1]

Capital Investment	10^3 \$	\$/kW	
Process Capital	33814.2	56.36	
General Facilities	1690.7	2.82	
Engineering and Home Office Fees	3381.4	5.64	
Project Contingency	6762.8	11.27	
Process Contingency	3381.4	5.64	
Total Plant Cost (Dec. 1985 dollars)	49030.6	81.72	
Total Plant Investment (Jan. 1986 startup)	49030.6	81.72	
Preproduction Costs	5140.7	8.57	
Inventory Capital	1660.0	2.77	
Total Capital Requirement (Jan. 1986 startup)	55831.3	93.05	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 \$	\$/kW-yr	
Operating Labor	307.1	0.51	
Maintenance Labor	541.0	0.90	
Maintenance Materials	811.5	1.35	
Administrative and Support Labor	254.4	0.42	
Total O&M First Year	1914.1	3.19	
Fixed O&M First Year	1244.2	2.07	
Variable O&M First Year	669.9	0.20	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 \$	mills/kW-hr	
Limestone	6473.9	1.89	
In-Plant Power (5124.3 KW)	1692.3	0.50	
Other Consumables: Water	8.9	0.00	
Waste Disposal	2533.4	0.74	
Total Consumables (excluding fuel)	10708.5	3.13	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	8547.8	14.25	\$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	21170.4	6.20	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	21170.4	213.35	\$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	2879.1	4.80	$/kW-yr
15-Year Levelized Variable O&M	1550.1	0.45	mills/kW-hr
15-Year Levelized Consumables O&M	24779.5	7.25	mills/kW-hr
15-Year Levelized Carrying Charges	8547.8	14.25	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	37756.5	11.05	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	37756.5	380.68	$/ton

Notes: (1) Coal Sulfur Content: 3.7%
Calcium/Sulfur Mole Ratio: 2:1
SO_2 Removed in Boiler: 50%
Overall SO_2 Removal: 90%

TABLE 38: ECONOMIC ANALYSIS OF
400 MWE NEW UNIT WITH LIMB AND SPRAY DRYER (HIGH-SULFUR)[1]

Capital Investment	$10^3 \$$	$/kW	
Process Capital	25870.1	64.68	
General Facilities	1293.5	3.23	
Engineering and Home Office Fees	2587.0	6.47	
Project Contingency	5174.0	12.94	
Process Contingency	2587.0	6.47	
Total Plant Cost (Dec. 1985 dollars)	37511.6	93.78	
Total Plant Investment (Jan. 1986 startup)	37511.6	93.78	
Preproduction Costs	3558.4	8.90	
Inventory Capital	1106.6	2.77	
Total Capital Requirement (Jan. 1986 startup)	42176.6	105.44	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	$10^3 \$$	$/kW-yr	
Operating Labor	307.1	0.77	
Maintenance Labor	413.9	1.03	
Maintenance Materials	620.9	1.55	
Administrative and Support Labor	216.3	0.54	
Total O&M First Year	1558.2	3.90	
Fixed O&M First Year	1102.8	2.53	
Variable O&M First Year	545.4	0.24	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	$10^3 \$$	mills/kW-hr	
Limestone	4315.9	1.89	
In-Plant Power (3688.3 KW)	1218.1	0.54	
Other Consumables: Water	6.8	0.00	
Waste Disposal	1688.9	0.74	
Total Consumables (excluding fuel)	7229.7	3.17	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	6457.2	16.14	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	15245.1	6.69	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	15245.1	230.57	$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	2343.6	5.86	$/kW-yr
15-Year Levelized Variable O&M	1263.1	0.55	mills/kW-hr
15-Year Levelized Consumables O&M	16729.5	7.35	mills/kW-hr
15-Year Levelized Carrying Charges	6457.2	16.14	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	26793.4	11.76	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	26793.4	405.22	$/ton

Notes: (1) Coal Sulfur Content: 3.7%
Calcium/Sulfur Mole Ratio: 2:1
SO_2 Removed in Boiler: 50%
Overall SO_2 Removal: 90%

TABLE 39: ECONOMIC ANALYSIS OF
200 MWE NEW UNIT WITH LIMB AND SPRAY DRYER (HIGH-SULFUR)[1]

Capital Investment	10^3 $	$/kW	
Process Capital	13760.8	68.80	
General Facilities	688.0	3.44	
Engineering and Home Office Fees	1376.1	6.88	
Project Contingency	2752.2	13.76	
Process Contingency	1376.1	6.88	
Total Plant Cost (Dec. 1985 dollars)	19953.2	99.77	
Total Plant Investment (Jan. 1986 startup)	19953.2	99.77	
Preproduction Costs	1821.1	9.11	
Inventory Capital	553.3	2.77	
Total Capital Requirement (Jan. 1986 startup)	22327.6	111.64	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 $	$/kW-yr	
Operating Labor	307.1	1.54	
Maintenance Labor	220.2	1.10	
Maintenance Materials	330.3	1.65	
Administrative and Support Labor	158.2	0.79	
Total O&M First Year	1015.7	5.08	
Fixed O&M First Year	660.2	3.30	
Variable O&M First Year	355.5	0.31	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 $	mills/kW-hr	
Limestone	2157.9	1.89	
In-Plant Power (1797.4 KW)	593.6	0.52	
Other Consumables: Water	3.1	0.00	
Waste Disposal	844.5	0.74	
Total Consumables (excluding fuel)	3599.1	3.16	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	3418.4	17.09	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	8033.2	7.05	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	8033.2	242.98	$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	1527.7	7.64	$/kW-yr
15-Year Levelized Variable O&M	822.6	0.72	mills/kW-hr
15-Year Levelized Consumables O&M	8328.3	7.31	mills/kW-hr
15-Year Levelized Carrying Charges	3418.4	17.09	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	14097.0	12.38	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	14097.0	426.40	$/ton

Notes: (1) Coal Sulfur Content: 3.7%
Calcium/Sulfur Mole Ratio: 2:1
SO_2 Removed in Boiler: 50%
Overall SO_2 Removal: 90%

TABLE 40: ECONOMIC ANALYSIS OF 600 MWE NEW UNIT WITH LIMB AND SPRAY DRYER (LOW-SULFUR)[1]

Capital Investment	10^3 \$	\$/kW	
Process Capital	15537.8	25.90	
General Facilities	776.9	1.29	
Engineering and Home Office Fees	1553.8	2.59	
Project Contingency	3107.6	5.18	
Process Contingency	1553.8	2.59	
Total Plant Cost (Dec. 1985 dollars)	22529.8	37.55	
Total Plant Investment (Jan. 1986 startup)	22529.8	37.55	
Preproduction Costs	3505.1	5.84	
Inventory Capital	426.0	0.63	
Total Capital Requirement (Jan. 1986 startup)	26460.9	44.10	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 \$	\$/kW-yr	
Operating Labor	307.1	0.51	
Maintenance Labor	248.6	0.41	
Maintenance Materials	372.9	0.62	
Administrative and Support Labor	166.7	0.28	
Total O&M First Year	1095.3	1.83	
Fixed O&M First Year	711.9	1.19	
Variable O&M First Year	383.4	0.11	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 \$	mills/kW-hr	
Limestone	1661.3	0.49	
In-Plant Power (1491.5 KW)	492.6	0.14	
Other Consumables: Water	3.6	0.00	
Waste Disposal	505.4	0.15	
Total Consumables (excluding fuel)	2662.9	0.78	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	4051.2	6.75	\$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	7809.4	2.29	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	7809.4	788.98	\$/ton

15-Year Levelized O&M Costs	10^3 \$		
15-Year Levelized Fixed O&M	1647.3	2.75	\$/kW-yr
15-Year Levelized Variable O&M	887.2	0.26	mills/kW-hr
15-Year Levelized Consumables O&M	6162.0	1.80	mills/kW-hr
15-Year Levelized Carrying Charges	4051.2	6.75	\$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	12747.7	3.73	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	12747.7	1287.90	\$/ton

Notes: (1)	
Coal Sulfur Content:	0.55%
Calcium/Sulfur Mole Ratio:	2:1
SO_2 Removed in Boiler:	50%
Overall SO_2 Removal:	90%

TABLE 41: ECONOMIC ANALYSIS OF
400 MWE NEW UNIT WITH LIMB AND SPRAY DRYER (LOW-SULFUR)[1]

Capital Investment	10^3 $	$/kW	
Process Capital	12856.3	32.14	
General Facilities	642.8	1.61	
Engineering and Home Office Fees	1285.6	3.21	
Project Contingency	2571.3	6.43	
Process Contingency	1285.6	3.21	
Total Plant Cost (Dec. 1985 dollars)	18641.6	46.60	
Total Plant Investment (Jan. 1986 startup)	18641.6	46.60	
Preproduction Costs	2430.4	6.08	
Inventory Capital	284.0	0.71	
Total Capital Requirement (Jan. 1986 startup)	21356.0	53.39	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 $	$/kW-yr	
Operating Labor	307.1	0.77	
Maintenance Labor	205.7	0.51	
Maintenance Materials	308.6	0.77	
Administrative and Support Labor	153.8	0.38	
Total O&M First Year	975.2	2.44	
Fixed O&M First Year	633.8	1.58	
Variable O&M First Year	341.3	0.15	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 $	mills/kW-hr	
Limestone	1107.5	0.49	
In-Plant Power (898.3 KW)	296.7	0.13	
Other Consumables: Water	2.8	0.00	
Waste Disposal	336.9	0.15	
Total Consumables (excluding fuel)	1743.9	0.77	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	3269.6	8.17	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	5988.7	2.63	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	5988.7	907.57	$/ton

	10^3 \$		
15-Year Levelized O&M Costs			
15-Year Levelized Fixed O&M	1466.6	3.67	\$/kW-yr
15-Year Levelized Variable O&M	789.8	0.35	mills/kW-hr
15-Year Levelized Consumables O&M	4035.4	1.77	mills/kW-hr
15-Year Levelized Carrying Charges	3269.6	8.17	\$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	9561.4	4.20	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	9561.4	1449.00	\$/ton

Notes: (1)	
Coal Sulfur Content:	0.55%
Calcium/Sulfur Mole Ratio:	4:1
SO_2 Removed in Boiler:	50%
Overall SO_2 Removal:	70%

TABLE 42: ECONOMIC ANALYSIS OF
200 MWE NEW UNIT WITH LIME AND SPRAY DRYER (LOW-SULFUR)[1]

Capital Investment	10^3 \$	\$/kW	
Process Capital	7352.5	36.76	
General Facilities	367.6	1.84	
Engineering and Home Office Fees	735.2	3.68	
Project Contingency	1470.5	7.35	
Process Contingency	735.2	3.68	
Total Plant Cost (Dec. 1985 dollars)	10661.1	53.31	
Total Plant Investment (Jan. 1986 startup)	10661.1	53.31	
Preproduction Costs	1259.7	6.30	
Inventory Capital	142.0	0.71	
Total Capital Requirement (Jan. 1986 startup)	12062.8	60.31	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 \$	\$/kW-yr	
Operating Labor	370.1	1.85	
Maintenance Labor	117.6	0.59	
Maintenance Materials	176.5	0.88	
Administrative and Support Labor	127.4	0.64	
Total O&M First Year	728.6	3.64	
Fixed O&M First Year	473.6	2.37	
Variable O&M First Year	255.0	0.22	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 \$	mills/kW-hr	
Limestone	553.8	0.49	
In-Plant Power (451.5 KW)	149.1	0.13	
Other Consumables: Water	1.3	0.00	
Waste Disposal	165.8	0.15	
Total Consumables (excluding fuel)	872.7	0.77	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	1846.8	9.23	\$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	3448.1	3.03	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	3448.1	1045.00	\$/ton

	10^3 \$		
15-Year Levelized O&M Costs			
15-Year Levelized Fixed O&M	1095.9	5.48	\$/kW-yr
15-Year Levelized Variable O&M	590.1	0.52	mills/kW-hr
15-Year Levelized Consumables O&M	2019.4	1.77	mills/kW-hr
15-Year Levelized Carrying Charges	1846.8	9.23	\$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	5552.2	4.88	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	5552.2	1682.69	\$/ton

Notes: (1) Coal Sulfur Content: 0.55%

Calcium/Sulfur Mole Ratio: 4:1

SO_2 Removed in Boiler: 50%

Overall SO_2 Removal: 70%

TABLE 43: ECONOMIC ANALYSIS OF 600 MWE NEW UNIT WITH FGD (HIGH-SULFUR)[1,2]

Capital Investment	10^3 $	$/kW	
Process Capital	86000.0	143.33	
General Facilities	4300.0	7.17	
Engineering and Home Office Fees	8600.0	14.33	
Project Contingency	17200.0	28.67	
Process Contingency	8600.0	14.33	
Total Plant Cost (Dec. 1985 dollars)	124700.0	207.83	
Total Plant Investment (Jan. 1986 startup)	124700.0	207.83	
Preproduction Costs	7165.4	11.94	
Inventory Capital	918.4	1.53	
Total Capital Requirement (Jan. 1986 startup)	132783.8	221.31	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 $	$/kW-yr	
Operating Labor	307.1	0.51	
Maintenance Labor	1376.0	2.29	
Maintenance Materials	2064.0	3.44	
Administrative and Support Labor	504.9	0.84	
Total O&M First Year	4252.0	7.09	
Fixed O&M First Year	2763.8	4.61	
Variable O&M First Year	1488.2	0.44	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 $	mills/kW-hr	
Limestone	3257.1	0.95	
In-Plant Power	4262.3	1.25	
Steam	3151.7	0.92	
Water	187.2	0.05	
Diesel Fuel	324.5	0.09	
Waste Disposal	2400.0	0.70	
Total Consumables (excluding fuel)	13582.8	3.98	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	20329.2	33.88	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	38164.0	11.17	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	38164.0	384.78	$/ton

	10^3 \$		
15-Year Levelized O&M Costs			
15-Year Levelized Fixed O&M	6395.4	10.66	\$/kW-yr
15-Year Levelized Variable O&M	3443.7	1.00	mills/kW-hr
15-Year Levelized Consumables O&M	31430.6	9.20	mills/kW-hr
15-Year Levelized Carrying Charges	20329.2	33.88	\$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	61598.8	18.03	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	61598.8	621.07	\$/ton

Notes:		
(1)	Coal Sulfur Content:	3.7%
	Calcium/Sulfur Mole Ratio:	--
	SO_2 Removed in Boiler:	--
	Overall SO_2 Removal:	90%
(2)	Limestone FGD with Forced Oxidation	

TABLE 44: ECONOMIC ANALYSIS OF 400 MWE NEW UNIT WITH FGD (HIGH-SULFUR)[1,2]

Capital Investment	$10^3\ \$$	\$/kW	
Process Capital	64000.0	160.00	
General Facilities	3200.0	8.00	
Engineering and Home Office Fees	6400.0	16.00	
Project Contingency	12800.0	32.00	
Process Contingency	6400.0	16.00	
Total Plant Cost (Dec. 1985 dollars)	92800.0	232.00	
Total Plant Investment (Jan. 1986 startup)	92800.0	232.00	
Preproduction Costs	5282.2	13.21	
Inventory Capital	619.8	1.55	
Total Capital Requirement (Jan. 1986 startup)	98702.0	246.76	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	$10^3\ \$$	\$/kW-yr	
Operating Labor	307.1	0.77	
Maintenance Labor	1024.0	2.56	
Maintenance Materials	1536.0	3.84	
Administrative and Support Labor	399.3	1.00	
Total O&M First Year	3266.4	8.17	
Fixed O&M First Year	2123.2	5.31	
Variable O&M First Year	1143.3	0.50	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	$10^3\ \$$	mills/kW-hr	
Limestone	2183.7	0.96	
In-Plant Power	2881.5	1.27	
Steam	2151.1	0.94	
Water	123.6	0.05	
Diesel Fuel	233.6	0.10	
Waste Disposal	1609.2	0.71	
Total Consumables (excluding fuel)	9182.7	4.03	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	15111.3	37.78	\$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	27560.4	12.10	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	27560.4	416.82	\$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	4913.1	12.28	$/kW-yr
15-Year Levelized Variable O&M	2645.6	1.16	mills/kW-hr
15-Year Levelized Consumables O&M	21248.8	9.33	mills/kW-hr
15-Year Levelized Carrying Charges	15111.3	37.78	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	43918.8	19.28	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	43918.8	664.22	$/ton

Notes: (1) Coal Sulfur Content: 3.7%
Calcium/Sulfur Mole Ratio: --
SO_2 Removed in Boiler: --
Overall SO_2 Removal: 90%

(2) Limestone FGD with Forced Oxidation

TABLE 45: ECONOMIC ANALYSIS OF 200 MWE NEW UNIT WITH FGD (HIGH-SULFUR)[1,2]

Capital Investment	$10^3 \$$	$/kW	
Process Capital	42000.0	210.00	
General Facilities	2100.0	10.50	
Engineering and Home Office Fees	4200.0	21.00	
Project Contingency	8400.0	42.00	
Process Contingency	4200.0	21.00	
Total Plant Cost (Dec. 1985 dollars)	60900.0	304.50	
Total Plant Investment (Jan. 1986 startup)	60900.0	304.50	
Preproduction Costs	2924.9	14.62	
Inventory Capital	322.6	1.61	
Total Capital Requirement (Jan. 1986 startup)	64147.5	320.74	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	$10^3 \$$	$/kW-yr	
Operating Labor	307.1	1.54	
Maintenance Labor	672.0	3.36	
Maintenance Materials	1008.0	5.04	
Administrative and Support Labor	293.7	1.47	
Total O&M First Year	2280.8	11.40	
Fixed O&M First Year	1482.5	0.91	
Variable O&M First Year	798.3	0.70	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	$10^3 \$$	mills/kW-hr	
Limestone	1128.9	0.99	
In-Plant Power	4262.3	3.74	
Steam	3151.7	2.77	
Water	187.2	0.16	
Diesel Fuel	324.5	0.28	
Waste Disposal	818.2	0.72	
Total Consumables (excluding fuel)	4802.0	4.22	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	9821.0	49.11	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	16903.8	14.84	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	16903.8	511.30	$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	3430.5	17.15	$/kW-yr
15-Year Levelized Variable O&M	1847.3	1.62	mills/kW-hr
15-Year Levelized Consumables O&M	11111.8	9.76	mills/kW-hr
15-Year Levelized Carrying Charges	9821.0	49.11	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	26210.6	23.02	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	26210.6	792.80	$/ton

Notes: (1) Coal Sulfur Content: 3.7%
Calcium/Sulfur Mole Ratio: --
SO_2 Removed in Boiler: --
Overall SO_2 Removal: 90%

(2) Limestone FGD with Forced Oxidation

TABLE 46: ECONOMIC ANALYSIS OF 600 MWE NEW UNIT WITH FGD (LOW-SULFUR)[1,2]

Capital Investment	10^3 $	$/kW	
Process Capital	59000.0	98.33	
General Facilities	2950.0	4.92	
Engineering and Home Office Fees	5900.0	9.83	
Project Contingency	11800.0	19.67	
Process Contingency	5900.0	9.83	
Total Plant Cost (Dec. 1985 dollars)	85550.0	142.58	
Total Plant Investment (Jan. 1986 startup)	85550.0	142.58	
Preproduction Costs	4985.3	8.31	
Inventory Capital	98.8	0.16	
Total Capital Requirement (Jan. 1986 startup)	90634.1	151.06	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 $	$/kW-yr	
Operating Labor	307.1	0.51	
Maintenance Labor	944.0	1.57	
Maintenance Materials	1416.0	2.36	
Administrative and Support Labor	375.3	0.63	
Total O&M First Year	3042.4	5.07	
Fixed O&M First Year	1977.6	3.30	
Variable O&M First Year	1064.9	0.31	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 $	mills/kW-hr	
Limestone	286.1	0.09	
In-Plant Power	2821.5	0.83	
Steam	0.00	0.00	
Water	119.5	0.03	
Diesel Fuel	96.1	0.03	
Waste Disposal	218.2	0.06	
Total Consumables (excluding fuel)	3551.4	1.04	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	13876.1	23.13	$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	20469.9	5.99	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	20469.9	2068.08	$/ton

	10^3 \$		
15-Year Levelized O&M Costs			
15-Year Levelized Fixed O&M	4576.2	7.63	\$/kW-yr
15-Year Levelized Variable O&M	2464.2	0.72	mills/kW-hr
15-Year Levelized Consumables O&M	8217.9	2.41	mills/kW-hr
15-Year Levelized Carrying Charges	13876.1	23.13	\$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	29134.4	8.53	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	29134.4	2943.46	\$/ton

Notes: (1) Coal Sulfur Content: 0.55%
Calcium/Sulfur Mole Ratio: --
SO_2 Removed in Boiler: --
Overall SO_2 Removal: 70%

(2) Limestone FGD with Forced Oxidation

TABLE 47: ECONOMIC ANALYSIS OF 400 MWE NEW UNIT WITH FGD (LOW-SULFUR)[1,2]

Capital Investment	10^3 \$	\$/kW	
Process Capital	45000.0	112.50	
General Facilities	2250.0	5.63	
Engineering and Home Office Fees	4500.0	11.25	
Project Contingency	9000.0	22.50	
Process Contingency	4500.0	11.25	
Total Plant Cost (Dec. 1985 dollars)	65250.0	163.13	
Total Plant Investment (Jan. 1986 startup)	65250.0	163.13	
Preproduction Costs	3793.1	9.48	
Inventory Capital	72.0	0.18	
Total Capital Requirement (Jan. 1986 startup)	69115.1	172.79	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	10^3 \$	\$/kW-yr	
Operating Labor	307.1	0.77	
Maintenance Labor	720.0	1.80	
Maintenance Materials	1080.0	2.70	
Administrative and Support Labor	308.1	0.77	
Total O&M First Year	2415.2	6.04	
Fixed O&M First Year	1569.9	3.92	
Variable O&M First Year	845.3	0.37	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	10^3 \$	mills/kW-hr	
Limestone	185.1	0.08	
In-Plant Power	1921.0	0.84	
Steam	0.00	0.00	
Water	79.7	0.03	
Diesel Fuel	96.1	0.04	
Waste Disposal	136.4	0.06	
Total Consumables (excluding fuel)	2418.2	1.06	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	10581.5	26.45	\$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	15414.9	6.77	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	15414.9	2336.09	\$/ton

	$10^3\ \$$		
15-Year Levelized O&M Costs			
15-Year Levelized Fixed O&M	3632.7	9.08	$/kW-yr
15-Year Levelized Variable O&M	1956.0	0.86	mills/kW-hr
15-Year Levelized Consumables O&M	5595.7	2.46	mills/kW-hr
15-Year Levelized Carrying Charges	10581.5	26.45	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	21765.9	9.56	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	21765.9	3298.56	$/ton

Notes: (1) Coal Sulfur Content: 0.55%
Calcium/Sulfur Mole Ratio: --
SO_2 Removed in Boiler: --
Overall SO_2 Removal: 70%

(2) Limestone FGD with Forced Oxidation

TABLE 48: ECONOMIC ANALYSIS OF 200 MWE NEW UNIT WITH FGD (LOW-SULFUR)[1,2]

Capital Investment	10^3 \$	\$/kW	
Process Capital	30000.0	150.00	
General Facilities	1500.0	7.50	
Engineering and Home Office Fees	3000.0	15.00	
Project Contingency	6000.0	30.00	
Process Contingency	3000.0	15.00	
Total Plant Cost (Dec. 1985 dollars)	43500.0	217.50	
Total Plant Investment (Jan. 1986 startup)	43500.0	217.50	
Preproduction Costs	2085.0	10.43	
Inventory Capital	48.4	0.24	
Total Capital Requirement (Jan. 1986 startup)	45633.8	228.17	
Operating and Maintenance Costs (First Year - Dec. 1985 dollars)	**10^3 \$**	**\$/kW-yr**	
Operating Labor	307.1	1.54	
Maintenance Labor	480.0	2.40	
Maintenance Materials	720.0	3.60	
Administrative and Support Labor	236.1	1.18	
Total O&M First Year	1743.2	8.72	
Fixed O&M First Year	1133.1	5.67	
Variable O&M First Year	610.1	0.54	mills/kW-hr
Consumables Operating Costs Excluding Fuel (First Year - Dec. 1985 dollars)	**10^3 \$**	**mills/kW-hr**	
Limestone	92.5	0.08	
In-Plant Power	1020.5	0.90	
Steam	0.0	0.00	
Water	39.8	0.03	
Diesel Fuel	96.1	0.08	
Waste Disposal	68.2	0.06	
Total Consumables (excluding fuel)	1317.2	1.16	
Fuel Cost (Increment) (First Year - Dec. 1985 dollars)	0.0	0.00	mills/kW-hr
Carrying Charges (First Year - Dec. 1985 dollars)	6986.5	34.93	\$/kW-yr
Busbar Cost of Power (First Year - Dec. 1985 dollars)	10046.9	8.82	mills/kW-hr
First Year Cost Effectiveness SO_2 Removed	10046.9	3044.88	\$/ton

15-Year Levelized O&M Costs	10^3 $		
15-Year Levelized Fixed O&M	2622.0	13.11	$/kW-yr
15-Year Levelized Variable O&M	1411.8	1.24	mills/kW-hr
15-Year Levelized Consumables O&M	3048.0	2.68	mills/kW-hr
15-Year Levelized Carrying Charges	6986.5	34.93	$/kW-yr
15-Year Levelized Busbar Cost of Power at Levelized Capacity Factor (0.65)	14068.3	12.35	mills/kW-hr
15-Year Levelized Cost Effectiveness SO_2 Removed	14068.3	4263.64	$/ton

Notes: (1) Coal Sulfur Content: 0.55%
Calcium/Sulfur Mole Ratio: --
SO_2 Removed in Boiler: --
Overall SO_2 Removal: 70%

(2) Limestone FGD with Forced Oxidation

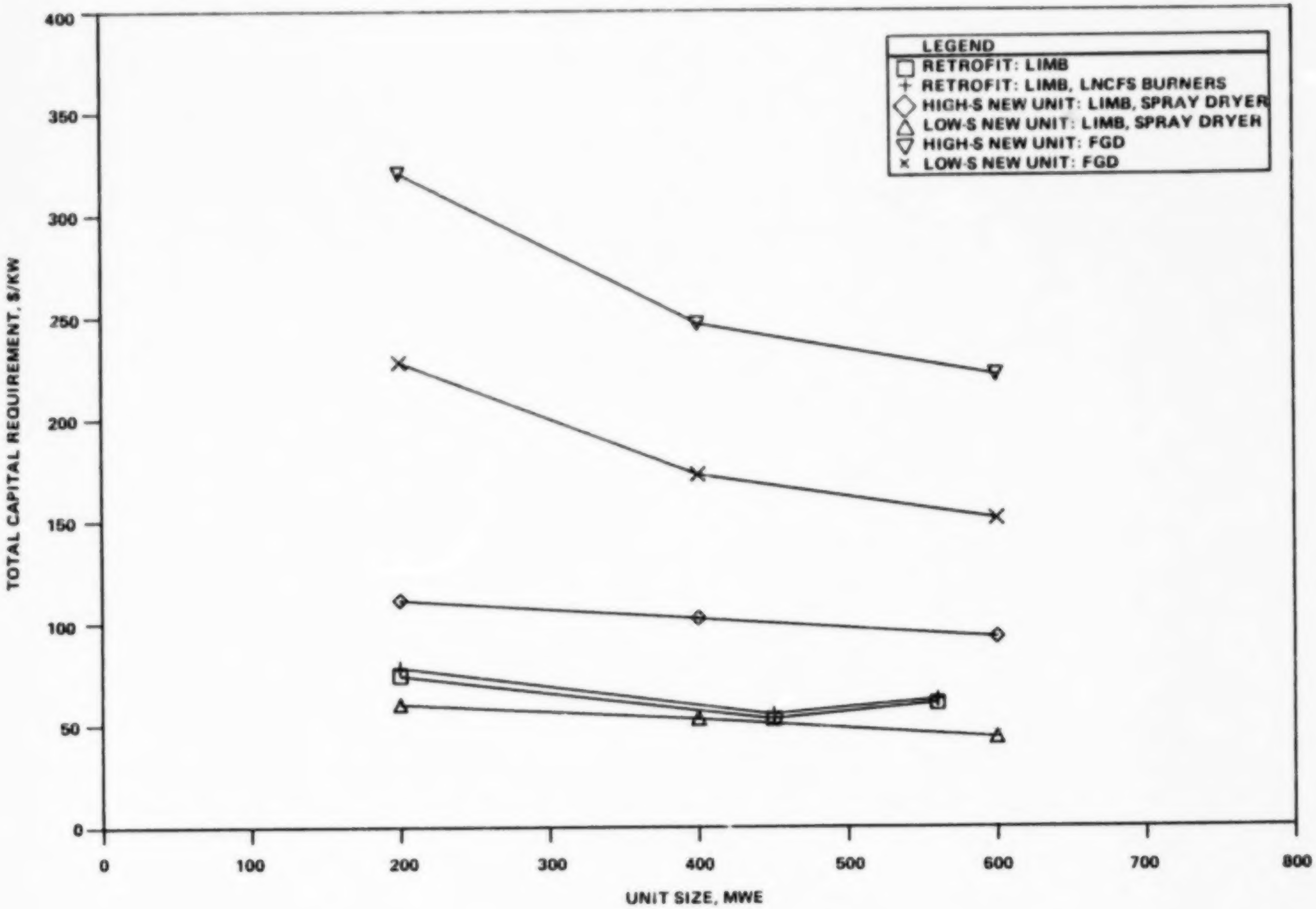


Figure 81. Total capital requirement versus unit size (January, 1986 startup).

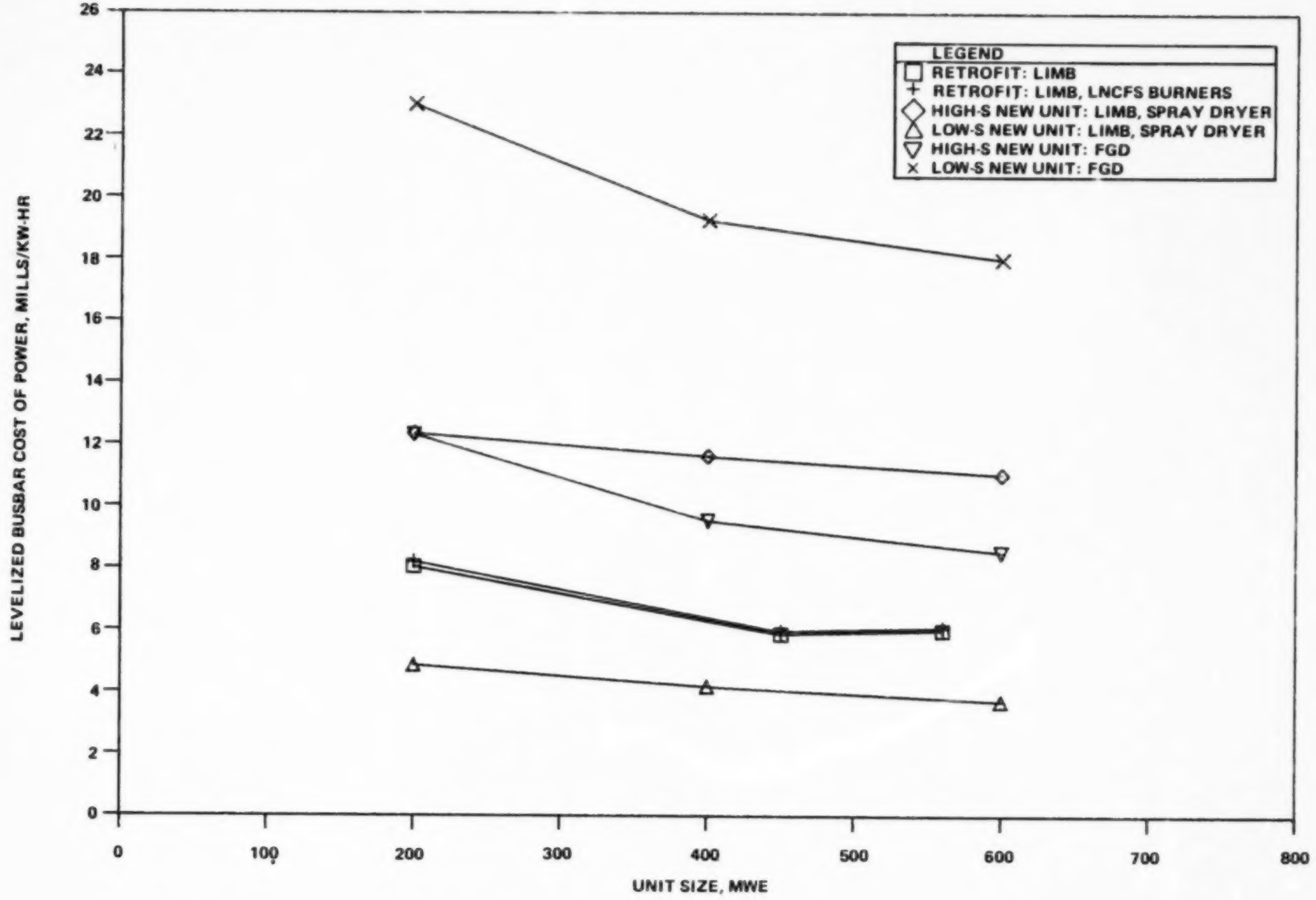


Figure 82. Levelized busbar cost of power versus unit size.

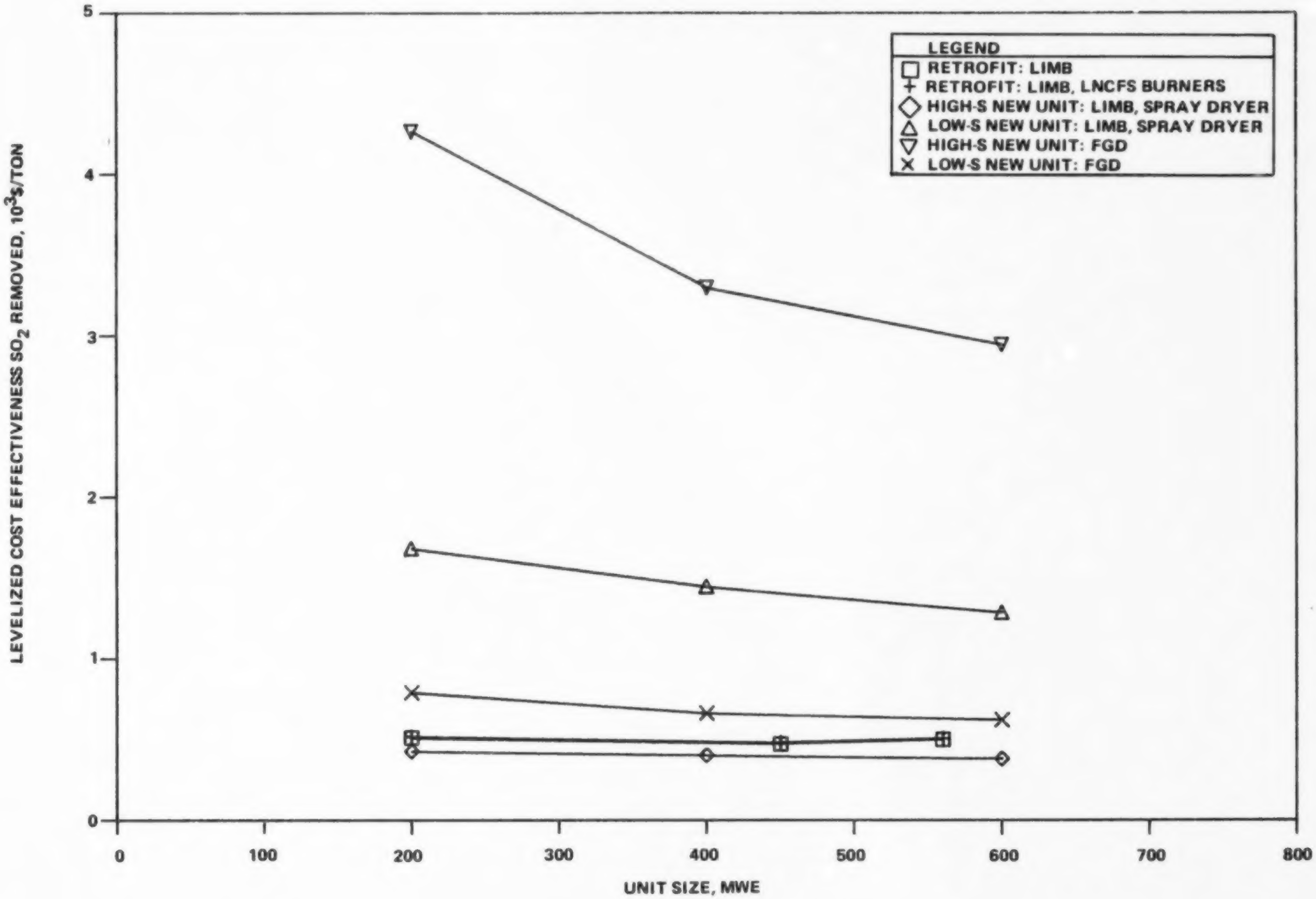


Figure 83. Levelized cost effectiveness SO_2 removed versus unit size.

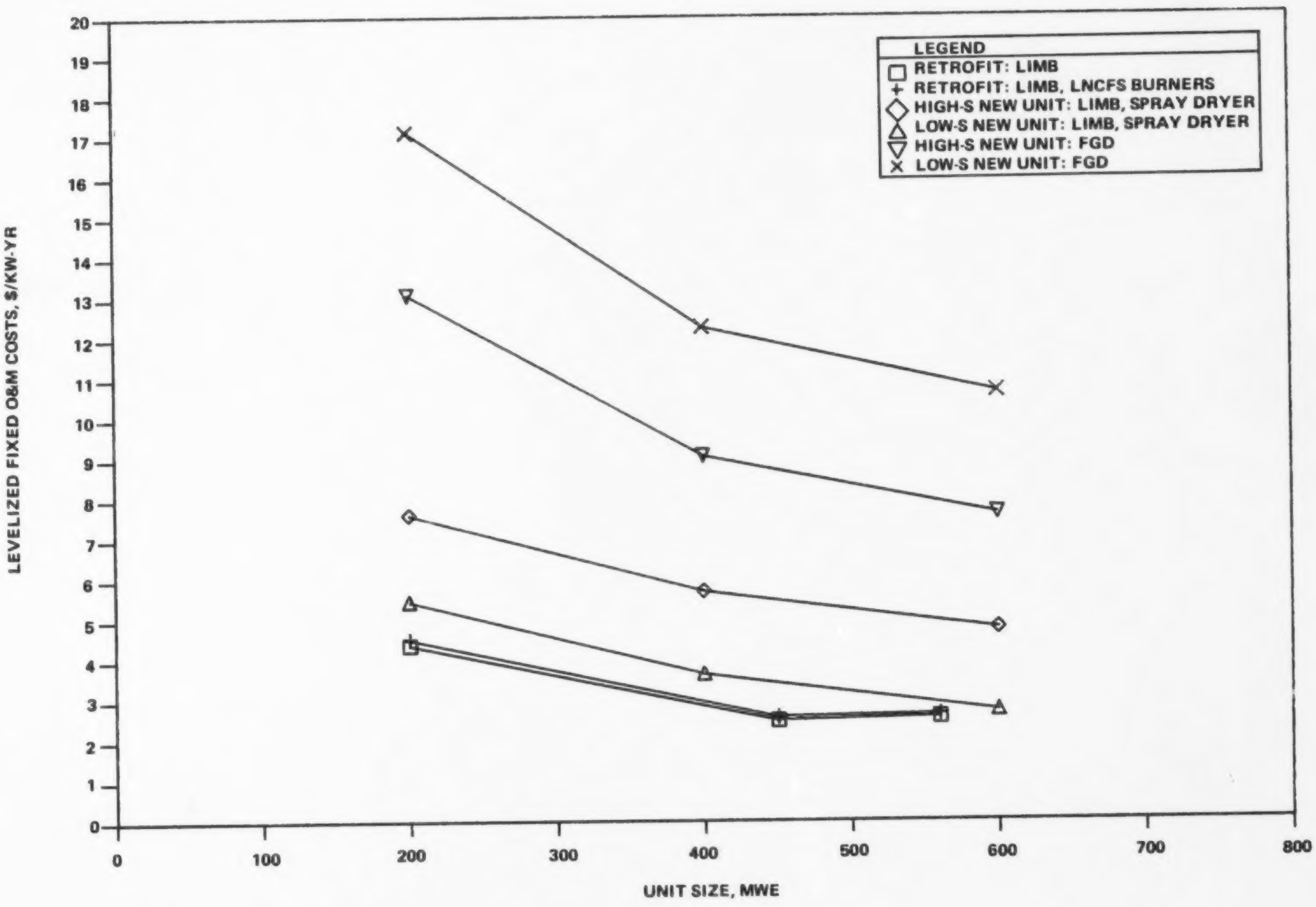


Figure 84. Levelized fixed O & M costs versus unit size.

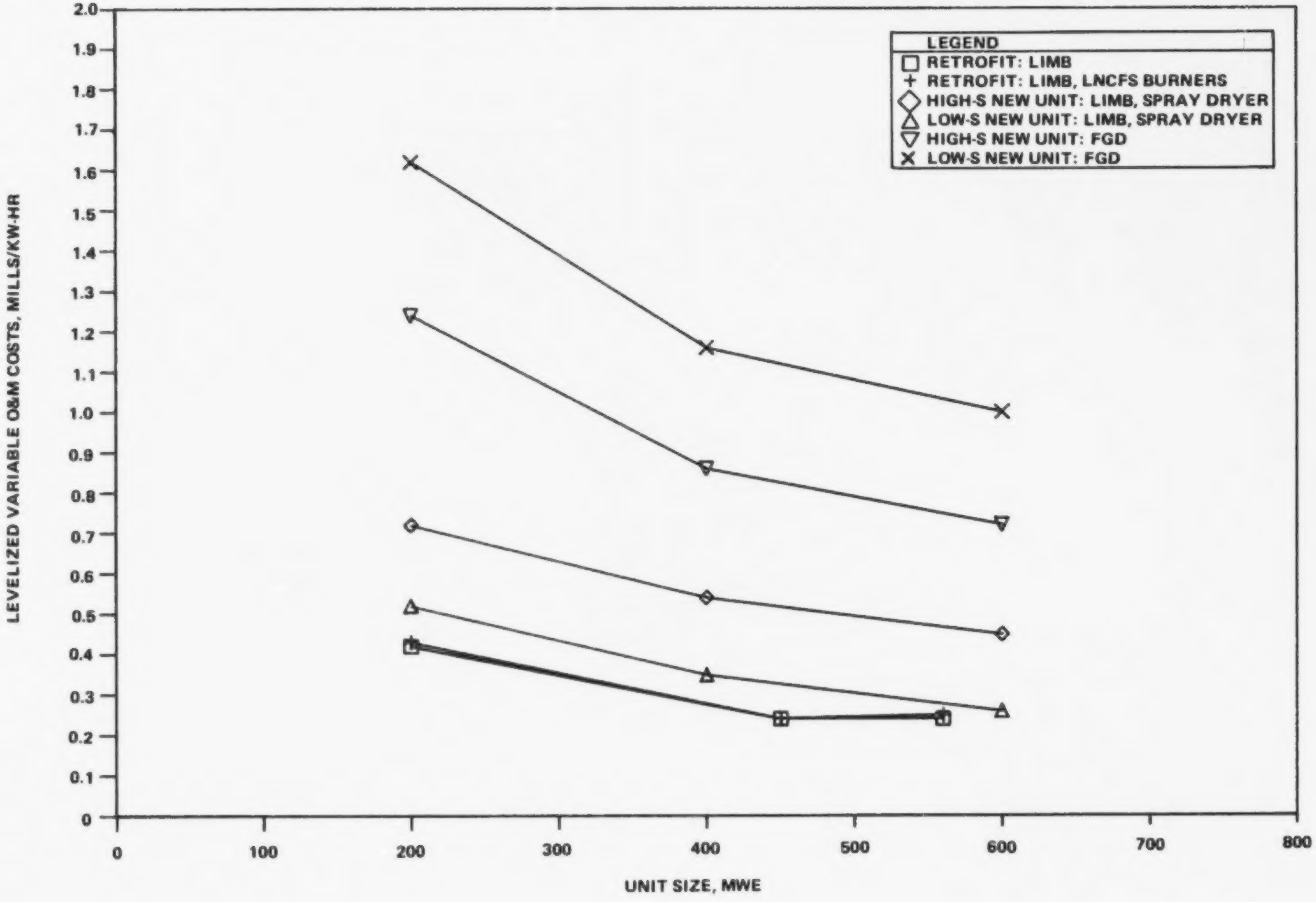


Figure 85. Levelized Variable O & M costs versus unit size.

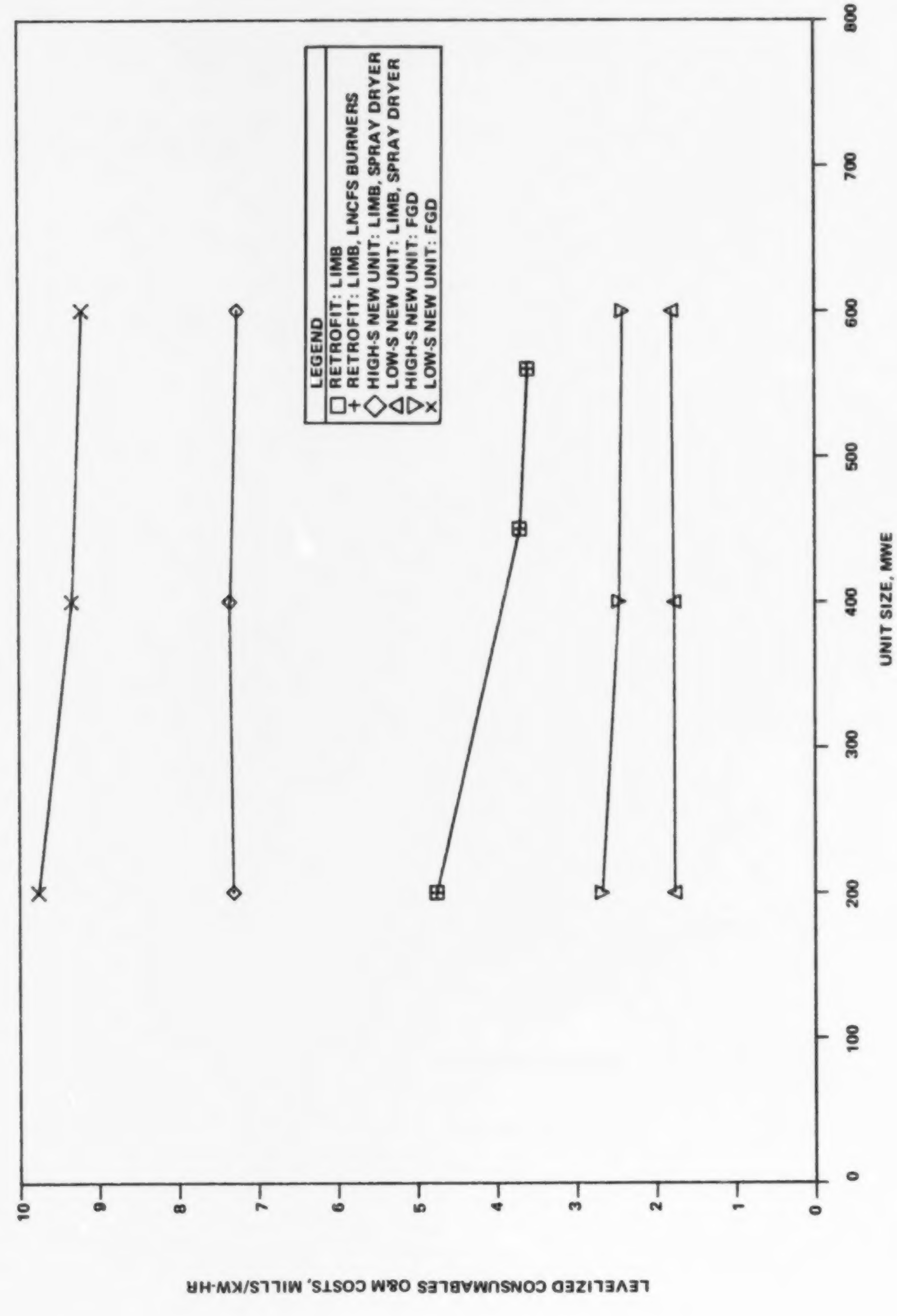


Figure 86. Levelized consumables O&M costs versus unit size.

TABLE 49. LIMB PROCESS CAPITAL COST BREAKDOWN[a]
(DECEMBER, 1985 DOLLARS)
(10^3 $)

Unit	Limestone Prep/Inj.	Boiler Modification	Back End Equipment	LNCFS Burners	Total[b]	Total[c]
Retrofit						
560 MWe	8947.8	4708.0	5047.1	860.0	18702.9	19562.9
450 MWe	7068.4	2481.0	4168.0	773.0	13717.4	14490.4
200 MWe	4665.8	1987.0	2012.9	607.0	8665.7	9272.7
New High-Sulfur						
600 MWe	12211.8	3690.0	17912.4	-	33814.2	-
400 MWe	8935.5	2667.0	14267.6	-	25870.1	-
200 MWe	4669.5	1531.0	7560.3	-	13760.8	-
New Low-Sulfur						
600 MWe	4016.3	1520.0	10001.5	-	15537.8	-
400 MWe	3217.5	1099.0	8539.8	-	12856.3	-
200 MWe	1829.7	633.0	4889.8	-	7352.5	-

[a] Additional cost breakdowns of specific systems are provided in Tables 50 through 53.

[b] No LNCFS burners.

[c] LNCFS burner upgrade included.

TABLE 50. LIMB PROCESS CAPTIAL COST BREAKDOWN
FOR SORBENT PREPARATION/INJECTION SYSTEM
MILL-RELATED COSTS
(DECEMBER, 1985 DOLLARS)
(10^3 \$)

Unit	Mills	Motors	Structural Steel	Enclosures	Valves	Mill Related Total	Overall System Total
Retrofit							
560 MWe	3867.0	187.1	632.0	151.2	22.5	4859.8	8947.8
450 MWe	3105.3	145.8	492.9	113.4	16.8	3874.2	7068.4
200 MWe	1933.5	93.5	352.5	75.6	11.2	2466.3	4665.8
New High-Sulfur							
600 MWe	5859.0	260.9	766.8	189.0	28.1	7103.8	12211.8
400 MWe	3867.0	187.1	632.0	151.2	22.5	4859.8	8935.5
200 MWe	1933.5	93.5	352.5	75.6	11.2	2466.3	4669.5
New Low-Sulfur							
600 MWe	1523.5	82.7	305.9	86.4	11.2	2009.7	4016.3
400 MWe	1078.1	22.9	225.9	86.4	11.2	1424.5	3217.5
200 MWe	539.2	11.4	147.3	43.2	5.6	746.7	1829.7

TABLE 51. LIMB PROCESS CAPTIAL COST BREAKDOWN FOR SORBENT PREPARATION/INJECTION SYSTEM NON-MILL-RELATED COSTS (DECEMBER, 1985 DOLLARS) (10^3 \$)

Unit	Fans	Bins	Structural Steel	Enclosures	Valves	Feeders & Elevators	Conveyors	Piping	Controls	Non-Mill-Related Total	Overall System Total
Retrofit											
560 MWe	95.9	167.8	1164.2	280.8	35.4	339.6	1070.5	642.2	291.6	4088.0	8947.8
450 MWe	73.9	132.8	896.4	210.6	26.6	254.7	802.9	504.7	291.6	3194.2	7068.4
200 MWe	48.0	83.8	629.9	140.4	17.7	169.8	535.2	283.1	291.6	2199.5	4665.8
New High-Sulfur											
600 MWe	129.2	244.9	1425.6	351.0	44.3	424.4	1338.1	858.9	291.6	5108.0	12211.8
400 MWe	95.9	167.8	1164.2	280.8	35.4	339.6	1070.5	629.9	291.6	4075.7	8935.5
200 MWe	48.0	83.8	629.9	140.4	17.7	169.8	535.2	286.8	291.6	2203.2	4669.5
New Low-Sulfur											
600 MWe	44.3	69.8	473.9	129.6	17.7	169.8	535.2	274.7	291.6	2006.6	4016.3
400 MWe	39.1	54.4	328.8	129.6	17.7	169.8	535.2	226.8	291.6	1793.0	3217.5
200 MWe	19.7	27.2	215.1	64.8	8.9	84.9	267.6	103.2	291.6	1083.0	1829.7

TABLE 52. LIMB PROCESS CAPTIAL COST BREAKDOWN FOR BOILER MODIFICATIONS (DECEMBER, 1985 DOLLARS) (10^3 $)

Unit	Surface Modifications Plus Penetrations[a]	Increased Sootblowing Capacity	Air Heater Upgrade	Total Boiler Modifications
Retrofit				
560 MWe	3832.0	876.0	-	4708.0
450 MWe	1798.5	682.5	-	2481.0
200 MWe	1248.5	738.5	-	1987.0
New High-Sulfur				
600 MWe	228.5	795.5	2666.0	3690.0
400 MWe	165.5	574.2	1927.3	2667.0
200 MWe	95.2	329.6	1106.2	1531.0
New Low-Sulfur				
600 MWe	477.5	1042.5	-	1520.0
400 MWe	344.5	754.5	-	1099.0
200 MWe	199.4	433.6	-	633.0

[a] Penetrations: retrofits only

TABLE 53. LIMB PROCESS CAPTIAL COST BREAKDOWN FOR BACK END EQUIPMENT (DECEMBER, 1985 DOLLARS) (10^3 \$)

Unit	Mechanical Collector	Ductwork	SO_3 Conditioning	Ash Handling Upgrade	Spray Dryer System	Baghouse[a]	Total
Retrofit							
560 MWe	1274.3	659.1	2706.9	406.8			5047.1
450 MWe	1065.6	553.3	2175.2	373.9			4168.0
200 MWe	536.1	289.2	966.7	220.9			2012.9
New High Sulfur							
600 MWe				513.4	17399.0	10559.0	17912.4
400 MWe				352.6	13915.0	14159.0	14267.6
200 MWe				207.3	7353.0	3970.0	7560.3
New Low Sulfur							
600 MWe				199.5	9802.0	9374.0	10001.5
400 MWe				129.8	8410.0	7318.0	8539.8
200 MWe				67.8	4822.0	3725.0	4889.8

[a] Baghouse cost not included in total.

associated ductwork, and SO_3 conditioning equipment. For new units, these costs include the spray dryers plus incremental ash handling capabilities over that which would be required without sorbent injection. The baghouse costs for the new units are also shown on Table 53, but, as these costs are not LIMB-related, they were not included in the process capital costs.

11.2.1 Retrofit Units

For the three units considered for this study:

- the cost of adding limestone injection ranged from 53.4 to 74.3 $/kW without modifying the burners. Adding the LNCFS low-NO_x burners as a retrofit option for NO_x reduction and potentially improved overall emissions performance added from 2.3 to 4.5 $/kW. The effect on cost of electricity of incorporating LIMB with LNCFS burners would range from 4.15 to 5.72 mills/kW-hr (15-year levelized cost ranged from 5.98 to 8.23 mills/kW-hr).

- incremental fuel requirements were negligible. Analysis of these units showed that design operating conditions could be met with only small losses in efficiency.

- incorporation of large multiclone collectors to remove a portion (from 56% to 71%) of the total entrained solids should reduce the cost of back end modifications relative to ESP upgrading.

11.2.2 New Units - High-Sulfur

For the three new high-sulfur units considered for this study:

- the cost of incorporating LIMB for 50% SO_2 removal plus a spray dryer/baghouse system to achieve 90% overall SO_2 removal ranged from 93.0 to 111.6 $/kW. The effect on cost of electricity ranged from 6.20 to 7.05 mills/kW-hr (30-year levelized cost ranged from 11.05 to 12.38 mills/kW-hr).

- incremental fuel charges were not considered. The effect of removing the potential for SO_3-related corrosion in the back end was to permit the lowering of the air heater outlet temperature 25°F to 255°F. This was accomplished in part by larger Ljungstrom air heaters which added to the process capital cost.

- the spray dryer costs represent non-standard arrangements based on all of the spray dryer sorbent material being taken from the baghouse as recycle. As was discussed in Section 7, this recycle approach may be 80% or more effective, relative to fresh lime, when the furnace calcined material passes directly to the spray dryer. A conventional spray dryer using only fresh lime with associated preparation equipment would cost 40% to 50% more for the same SO_2 removal effectiveness. The baghouse was not included in costing the LIMB-related costs, since its size would be roughly the same with or without LIMB.

- the cost of conventional flue gas desulfurization equipment was obtained by backing out the inconsistent costs, such as interest during construction, from the costs presented in Ref. (29) for limestone scrubbers with oxidation. Adjustments were made for sulfur content and unit size. The cost for high-sulfur units ranged from 221.3 to 320.7 \$/kW. The effect on cost of electricity was 11.17 to 14.84 mills/kW-hr (30-year levelized costs ranged from 18.03 to 23.02 mills/kW-hr).

11.2.3 New Units - Low-Sulfur

For the three low-sulfur units considered for this study:

- the cost of incorporating LIMB for 50% SO_2 removal plus a spray dryer/baghouse system to achieve 70% overall SO_2 removal ranged from 44.1 to 60.3 \$/kW. The effect on cost of electricity ranged from 2.29 to 3.03 mills/kW-hr (30-year levelized cost ranged from 3.73 to 4.88 mills/kW-hr).

- incremental fuel charges were not considered. While no changes were made to the air heater exit temperature, which was already at 255°F, design operating conditions were maintained by rearranging surfaces in the convective pass.

- the discussion of the spray dryer costs for the high-sulfur units which appears in the previous subsection also applies to the low-sulfur units. The major difference in the low sulfur spray dryer system was that 50% of the flue gas leaving the boiler was bypassed around the spray dryer and reintroduced upstream of the baghouse. Costs for associated ductwork required for the bypass are included.

- the cost of conventional flue gas desulfurization equipment was obtained in a similar manner to that used for high-sulfur coal units as discussed in the previous subsection. The cost for low-sulfur units ranged from 151.1 to 228.2 $/kW. The effect on cost of electricity was 5.99 to 8.82 mills/kW-hr (30-year levelized costs ranged from 8.53 to 12.35 mills/kW-hr).

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